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EVALUATION AND UPGRADING OF A MULTI-STAGE TRICKLING FILTER FACILITY

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FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory - Cincinnati (IERL-CI) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report evaluates the applicability of several alternate treatment methods for complex organic waste waters and documents the results of an in-depth study of a large scale, multi-stage trickling filter system. From the study, it was concluded that the anticipated acclimation of the microorganisms and enhanced BOD removal were not achievable but rather, that undesirable air stripping of volatile organic constituents contributed very significantly to apparent pollutant removal.

This study is one of several undertaken by IERL-CI to evaluate alternate systems for treatment of complex wastes and reduction of discharges to our Nation's waterways. The report alerts industrial waste system designers, state officials, and EPA officials responsible for standard-setting to some of the problems which must be anticipated even with relatively conventional technology.

For further information on this subject, contact the Industrial Pollution Control Division, Organic Chemicals and Products Branch.

David G. Stephan
Director
Industrial Environmental Research Laboratory
Cincinnati

ABSTRACT

The applicability of a full-scale, six-stage trickling filter plant was investigated for the treatment of waste from a multiproduct organic chemical plant. Reductions of BOD per unit volume of packing indicated that the series design of the system did not lead to stagewise acclimation of the microorganisms or enhanced BOD removals. Tests to determine BOD removal mechanisms in the system indicated that air stripping and biological mechanisms both contributed significantly to the total observed reduction.

Effluent recycle considerably improved filter performance. However, 600 percent recycle was required for an approximate 90 percent reduction of BOD at a hydraulic loading of 2 gpm/sq ft (0.08 cu m/min-sq m).

Bench-scale activated sludge investigations showed that this process could be used successfully for upgrading the trickling filter system. Kinetic parameters necessary for the design of activated sludge systems were determined. Comparative studies of air and oxygen aeration indicated that the treatment of wastes containing volatile substances might be difficult in a closed oxygen system.

Activated carbon adsorption, also tested in small-scale systems, was most effective as a means of upgrading the trickling filter system when applied following activated sludge treatment. Activated carbon tests indicated that some refractory organics and color-producing substances could be effectively removed by adsorption; however, a significant nonadsorbable organic fraction remained.

This report was submitted by AWARE, Inc. in fulfillment of U. S. Environmental Protection Agency Grant 12020 FOH to CIBA-GEIGY Corporation.

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SECTION I

CONCLUSIONS

1. Over the period of this investigation, the six-stage trickling filter system removed only 32 percent of the effluent BOD load. The effluent BOD averaged 2,180 mg/l. Due to extreme organic overloading in the system, no stage-wise acclimation of microorganisms was observed. The quantity of BOD removed in the system ranged from approximately 100 to 200 lb/1,000 cu ft-day (1.5 to 3 kg/cu m). Since removals of this magnitude are not uncommon for single-stage, plastic medium trickling filter systems, it cannot be concluded that multiple staging resulted in extraordinarily high removals/unit volume for this waste.

2. Recycle investigations in three stages of the full-scale system demonstrated that approximately 50 percent BOD removal could be achieved at 20°C using a recirculation ratio of 2.3. However, while recirculation improved filter performance, these results and small-scale trickling filter tests indicated that BOD reductions higher than approximately 90 percent could not be achieved at recycle ratios as high as 800 percent. For this high strength waste (initial BOD of 3,200 mg/l), this level of treatment would probably be inadequate in most instances.

3. The six-stage trickling filter system, even when modified to include effluent recycle, failed to produce acceptable results as the sole treatment method for the wastewater. Due to the failure of the system to provide extraordinarily high organic removals, their multi-stage design can be used only as an initial treatment step.

4. Investigations made using the small-scale trickling filter to determine the mechanism of BOD removal indicated the principal removal mechanism at low recycle ratios was by air stripping, but that the fraction of BOD removed by air stripping decreased significantly as the recycle rate was increased.

5. The annual treatment cost for the 1.44 mgd facility including neutralization, equalization, trickling filtration in a six-stage system, and clarification totaled \$690,000/yr including capital and operating and maintenance costs. This was equivalent to \$1.58/1,000 gal and \$0.19/lb BOD removed.

6. Bench-scale activated sludge investigations showed that this process could be used to upgrade the trickling filter effluent. Treatment of the waste was characterized by a relatively high oxygen demand per unit of BOD removed, a relatively low sludge yield, and poor oxygen transfer characteristics. The effluent was characterized by high effluent suspended solids concentrations ranging from 90 to 175 mg/l and a high refractory effluent COD of approximately 450 mg/l.

7. Performance of the activated sludge process was significantly impaired using pure oxygen aeration in a closed atmosphere system treating the unstripped raw wastewater. However, performance using aeration treating a stripped waste was equal to or better than that obtained with air aeration.

8. Activated carbon adsorption was capable of removing refractory organic substances from the activated sludge effluent. Results indicated that treatment of the trickling filter effluent using this process would be feasible following removal of high concentrations of short-chained solvents.

9. Results of air stripping investigations showed that a significant fraction of the raw waste organics could be removed in an aerated equalization basin with surface aerators and in packed towers such as trickling filter systems. In order to achieve 60 percent BOD reductions, an air flow of 500 cu ft/gal was required in the packed tower, while a 3-day detention time using 150 hp/mil gal aeration capacity was required in the basin. These results indicate that the effects of stripping especially as they relate to scale-up from small-scale systems can be significant in designing biological systems to treat wastes containing volatile substances.

SECTION II

RECOMMENDATIONS

1. The results of this investigation indicated that waste produced in the manufacture of miscellaneous organic chemicals can be treated using both biological and physical-chemical processes. However, the complexity of the waste and large fluctuations in waste composition caused by changes in production schedules dictate that long-term responses of applicable treatment systems be thoroughly considered prior to preparing a final design. Primary consideration should be given to designs in which fluctuations in concentration and composition have a minimum effect on system performance.

2. In designing plastic medium trickling filters it is important that consideration be given to the structural stability of the packing and additional requirements for vertical and lateral supports in the tower superstructure. Packing such as the one used in this design has a minimum ability to support vertical and lateral loads and, therefore, must be strengthened by more complex supporting structures than would be required for other commercially available packing designs.

3. Effluent recirculation should be included in trickling filters designed for treatment of high-strength industrial wastes.

4. In designing multi-stage trickling filters, care should be taken not to overload the first stages of a system to the extent that failure of the entire system will be experienced.

SECTION III

INTRODUCTION

Wastes from many organic chemical plants typically contain significant concentrations of short-chained organics used as solvents and lesser concentrations of more complex organic substances. In comparison to municipal sewage, these wastes are highly variable in composition depending on plant production schedules, are composed principally of soluble organic substances which cannot be removed by precipitation or adsorption onto biological flocs, and may contain exotic chemical species which would interfere with conventional treatment processes. When treating wastes of this type, biological processes would be most effective in removing short-chained and other degradable organic constituents, while physical and chemical processes, such as adsorption using activated carbon, would be required to remove substances refractory to biological treatment.

The purpose of this project was to evaluate the use of multi-stage trickling filters and other methods for treating wastes produced in a diversified organic chemicals plant. The project involved demonstration of a full-scale, six-stage plastic medium trickling filter for reduction of organic constituents in the wastestream and the investigation of other potential treatment methods. Advantages which might be attributed to the multi-stage trickling filter include the sequential development of micro-organisms in the system acclimated to the different organic constituents in the waste. If this were the case, greater removals of organic constituents per unit volume might be experienced compared to conventional trickling filter designs. Small-scale tests using the activated sludge and activated carbon processes were performed to determine the applicability of these methods for further treatment of the waste required to meet anticipated best practicable and best available treatment technology.

BACKGROUND INFORMATION

The CIBA-GEIGY plant located at Cranston, Rhode Island is a multi-product organic chemicals plant in which over 500 different raw materials are used to make more than 200 different chemical products. Products manufactured in the plant include plastic additives, pharmaceuticals, dyestuffs and intermediates, and miscellaneous chemicals. The products produced here are destined for use largely in the plastics, leather, textiles and paper industries. Because of the diversity of products produced in the plant, it is difficult to precisely relate the plant production to the nature of the wastewaters produced. During 1973, 30,000,000 lb of products were manufactured in the plant. The production was distributed among various product categories as follows: pharmaceuticals, 3 percent; plastic additives such as antioxidants, 59 percent; chemical products (mostly fluorescent whitening agents and textile auxiliaries), 26 percent; tanning agents, 11 percent; and printing inks, 1 percent.

Situated on the Pawtuxet River approximately one mile upstream from Narragansett Bay, the plant is known for its modern facilities and versatility. Due to the extensive development and pilot plant facilities located at this site and the diverse nature of products manufactured in the plant, it has become known within CIBA-GEIGY Corporation as the "pilot plant of the world."

Originally, wastes generated in the plant were discharged through numerous outfalls to the Pawtuxet River. When it was recognized that direct disposal of untreated wastes to the river was unacceptable, a long-term wastewater management program was implemented in the plant. In 1965 cooling water used in the plant was separated from process discharges and combined in a recirculated cooling system. Approximately 7 mgd (26,000 cu m/day) of cooling water is recirculated through the plant.

In order to reduce process waste discharges, certain solvents and acids were segregated and either cleaned, concentrated and recycled or transported away from the plant for proper disposal. Having isolated most materials which lent themselves to this type of treatment, it was realized that additional treatment of wastes would be required. Investigations were initiated in July, 1965 to evaluate alternative means for treating and disposing of wastes in the plant. After considerable study a decision was made to construct a multi-stage plastic medium trickling filter system. This system was designed and constructed by Biodize Systems, Enviro-Chem Division of Monsanto Company and was placed in operation during the spring of 1972.

In 1970 CIBA-GEIGY Corporation received a research and development grant from the Environmental Protection Agency to evaluate the multi-stage trickling filter system over a two-and-one-half year period. Following start-up of the system, CIBA-GEIGY Corporation retained Associated Water and Air Resources Engineers, Inc. (AWARE) to perform the on-site investigations associated with the demonstration grant. In November, 1972 investigations were initiated.

THE APPLICATION OF MULTI-STAGE TRICKLING FILTERS

Although most work relating to pilot- and full-scale trickling filter installations has been concerned with plants in which only a single stage is employed, two- and three-stage systems are not uncommon. Advantages of multi-stage systems which have been advanced in the past include: greater flexibility in treating high-strength wastes to levels acceptable for discharge and the development of varied microbial populations from one stage to the next which might provide greater organic removals compared to a single stage having an equivalent volume. The primary disadvantage of multi-stage filters is the tendency to organically overload the first stages of the system.

Good performance of plastic medium, multiple-stage trickling filters has been reported for numerous industrial effluents (1, 2). Food processing industries routinely use multiple-stage plastic filters in the treatment of wastes from dairies, meat packing plants, fruit and vegetable processing plants, potato processing plants, and breweries (3, 4). In one case, a four-stage system was designed to treat a very strong pharmaceutical effluent (3, 5). The design included a series of four plastic filters, each with individual recycle and an individual settling tank (although interstage clarification is not considered essential) followed by a fifth stage consisting of a six-foot granite polishing filter. Recirculation ranging from 100 to 200 percent was practiced within each stage.

The effect of effluent recirculation cannot be predicted _a priori_ in any single case. In certain cases, negative effects have been reported. For the case of concentrated industrial effluents, however, most researchers have reported beneficial effects of recirculation (6, 7, 8). This is due to enrichment of specific enzymes, aeration, adsorption of inhibitory compounds, dilution, and dampening of short-term raw waste load variability.

Equally opposed opinions have been expressed concerning the existence of a limiting organic loading. The limiting loading is defined as a load beyond which the BOD removal decreases. The existence of such a loading has been considered by numerous investigators (9-13). Other authors, working primarily with plastic media, reported no limiting loading (14, 15).

OBJECTIVES

The principal objectives of this investigation were to evaluate a full-scale multi-stage plastic medium trickling filter system for the treatment of waste from a diversified organic chemical plant and to determine the applicability of the activated sludge and activated carbon processes for further treatment of the waste as judged by tests in small-scale systems. Specific objectives outlined in the grant application were modified during the course of the investigation by joint agreement between CIBA-GEIGY Corporation and the U.S. EPA. The restructured objectives of the investigation were as follows:

1. To optimize operation of present treatment facilities.
2. To examine operation of the multi-stage trickling filter system at various organic loads, temperature conditions, and other operating variables.
3. To investigate BOD removal mechanisms in the trickling filter system and to determine whether stage-wise acclimation of microorganisms occurred.
4. To evaluate the feasibility of using automatic monitoring equipment for the control of trickling filter systems.
5. To examine the activated sludge process as a means of further treating the waste and to determine the kinetic parameters required for a process design.
6. To examine activated carbon adsorption as a means of reducing BOD and removing refractory organic materials.

Personnel from AWARE, Inc. were responsible for sampling, operation and care of experimental treatment units and compilation and presentation of all data obtained from the project. CIBA-GEIGY personnel were responsible for the conduct of all analytical work, the operation and maintenance of automatic monitoring equipment and the conduct of effluent recycle investigations in the full-scale trickling filter system.

SECTION IV

DESCRIPTION OF TREATMENT FACILITY

Although this project was concerned primarily with the performance of the multi-stage trickling filter system, several other unit operations were required for treatment of the complex chemical waste discharged. The several unit operations included in the entire treatment system and characteristics of the influent waste are discussed in this chapter.

SOURCES OF WASTEWATER

The complexity of waste treatment problems at the Cranston plant is indicated by the large number of products produced in the facility and by the variety of manufacturing processes employed. Production of the great number of products required from this plant necessitates the use of batch processing methods which usually generate greater quantities of waste than comparable continuous-flow production methods. Since each reactor or "kettle" is used for producing a large number of products, segregation of wastes would be impossible. It is often common for products to be manufactured during "campaigns" lasting several weeks during which several months' supply of a particular product is produced. Therefore, in addition to daily fluctuations in wastewater strength and composition, significant weekly and longer-term variations are also evident.

The manufacturing plant consists of six buildings containing over 100 independent reactor units varying in size up to 3,000 gal (12,000 liters). In addition to agitated vessels, reaction units include appropriate auxiliary equipment such as distillation columns, condensers, receiving vessels, many types of filters and dryers and associated material transfer equipment.

Because of the many different reaction cycle times, no fixed number of batches per day can be quoted. However, the plant operates on a 24 hr/day basis and material transfer operations are in progress simultaneously on anywhere from 5 to 25 percent of the equipment.

Sources of wastewater include: 1) detergents, solvents, and other materials used to clean kettles, 2) solvents and other materials used to wash and purify products, 3) filtrates and centrates from product dewatering and drying, 4) water from vacuum jets, 5) blowdown from a recirculating cooling system, 6) still bottoms, 7) materials used to clean floors in production areas, and 8) stormwater run-off from the production area. Wastewater from manufacturing areas is collected in the plant sewer system and conveyed to the main pumping station. Based on the use of city water recorded in the production area and the quantity of wastewater received at the treatment plant, there is very little water loss in the plant due to evaporation or packaging with products. Comparison of these flows over the duration of the study showed that recordings from the magnetic flow meter in the treatment plant were approximately 15 percent greater than water meter readings. Some difference would be expected due to the introduction of stormwater flows to the plant sewer.

During most of the study period, work in the manufacturing areas proceeded 24-hr/day, 6 days/wk. However, at times the work schedule was decreased to a 5-day week or increased to 7 days/wk depending upon inventory backlogs. During periods when the plant was not operating, the waste flow averaged approximately 200 gpm (0.8 cu m/min) due to water used for vacuum jets and cooling systems.

WASTEWATER CHARACTERIZATION

Twenty-four hour composite samples of wastewater were collected over the nine month period from November 1972 through July 1973 and analyzed for BOD, COD, TSC (total carbon test performed on filtered wastewater, total soluble carbon), suspended solids, volatile suspended solids, color and temperature. The results of these and other tests performed on the wastewater are summarized in Table 1. The waste contained a high concentration of organic material, demonstrated by a total influent BOD of 3,190 mg/l and a total influent COD of 4,676 mg/l. The BOD/COD ratio of the raw waste was 0.68. At median concentrations approximately 90 percent of the total BOD and COD was in the soluble fraction, due largely to solvents present in the wastewater. Weekly grab samples were analyzed to determine the concentrations of various solvents. The ranges and average values for these solvents are shown in Table 2.

In order to determine the approximate fraction of the total organic waste load due to solvents, the BOD attributable to each solvent

TABLE 1

WASTEWATER CHARACTERIZATION

Parameter	Average	Percentile Values 10%	50%	90%
Flow[a] (mgd), all days	1.26	0.48	1.42	1.64
Weekdays	1.48	1.12	1.54	1.68
BOD (mg/l) Total	3190	1400	3000	5050
Soluble	3090	1200	2600	4750
COD (mg/l) Total	4676	2400	4650	6900
Soluble	4100	2100	4150	6200
TSC (mg/l)	1200	620	1220	1830
Suspended solids (mg/l)	151	48	108	288
Volatile suspended solids (mg/l)	131	44	100	250
Total kjeldahl nitrogen (mg/l)	37	2.5	30	94
Total phosphorus (mg/l)	6	2.2	3.8	13
Alkalinity (mg/l as $CaCO_3$ to pH 3.7)	227	155[b]	---	282[b]
Total solids (mg/l)	2725	1500[b]	---	4860[b]
Color (APHA units)	1800	400[b]	---	4000[b]
Temperature (°F) Influent	---	57	67	77
Effluent	---	52	62	72

[a] Flow in cu m/day = mgd x 3790.

[b] Values are minimum and maximum observed.

TABLE 2

SOLVENT DETERMINATIONS FOR RAW EQUALIZED WASTE

Solvent	Concentration (mg/l)	
	Average	Range
Acetone	280	30-600
Methyl ethyl ketone	20	10-100
Methanol	220	0-700
Isopropyl alcohol	1,260	500-3,100
Toluene	120	0-300

was calculated and correlated to the total BOD of the influent waste as shown in Figure 1. This correlation indicates that approximately 61 percent of the BOD was due to solvents. In a similar manner total carbon values were calculated for each solvent analysis. The results shown in Figure 2 indicate that practically all of the total carbon may be attributed to the solvent load. However, while total carbon values for the solvents were theoretical, less than 100 percent total carbon can be measured by any laboratory procedure used for the influent waste. Relationships were developed between total BOD and COD and between soluble BOD and TSC, resulting in the following equations:

$$BOD_T = 0.788 (COD_T) - 420$$

$$BOD_S = 2.65 (TSC) - 382$$

The suspended solids content of the waste averaged 151 mg/l; however, a significant fraction of the solids was very highly dispersed and non-settleable. Volatile suspended solids comprised 87 percent of the total suspended solids. Nitrogen and phosphorus concentrations of the raw wastewater were determined from composite samples collected from mid-November to mid-December of 1972. The quantities present would be insufficient

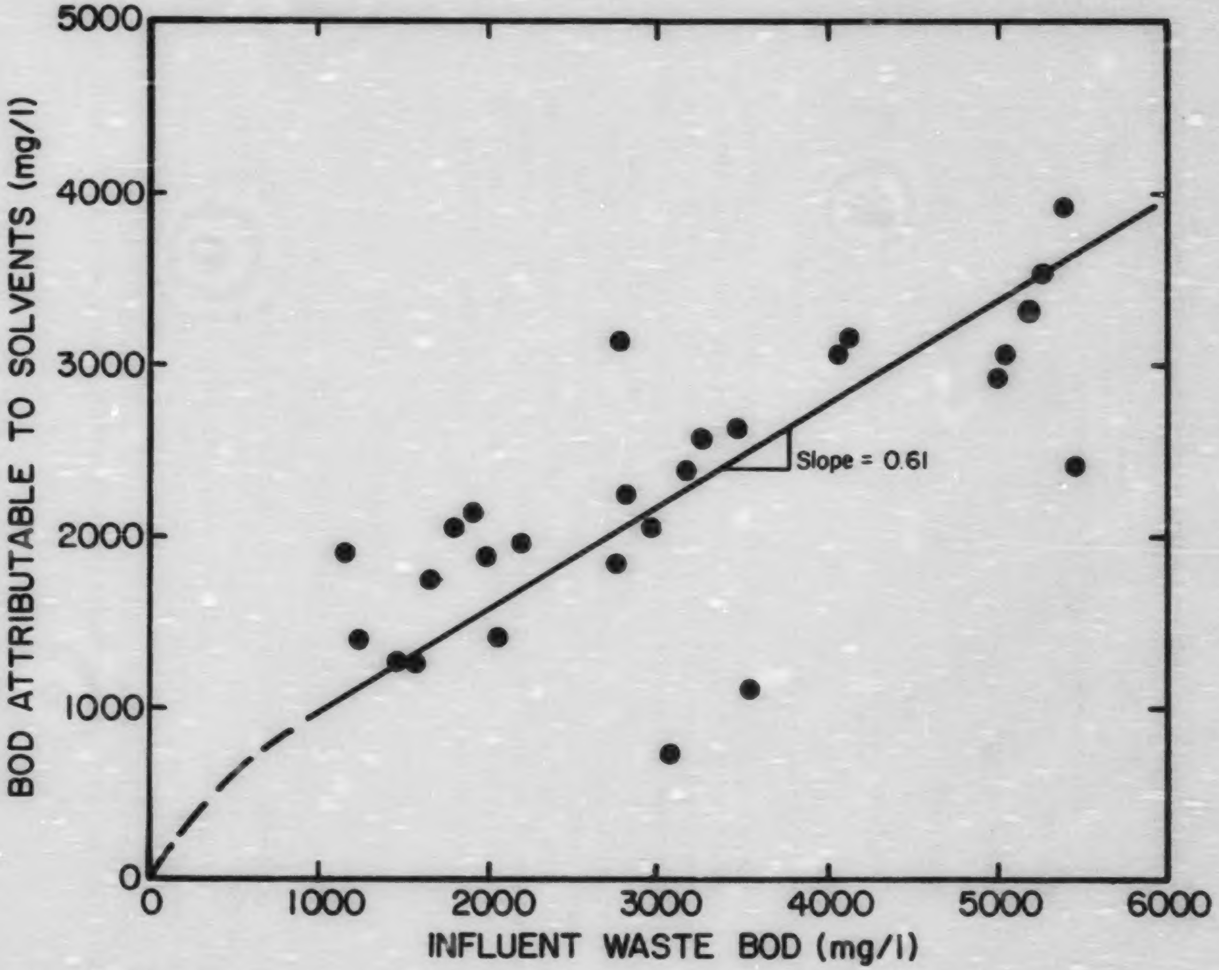


FIG. 1 . CORRELATION OF MEASURED BOD TO BOD ATTRIBUTABLE TO SOLVENTS

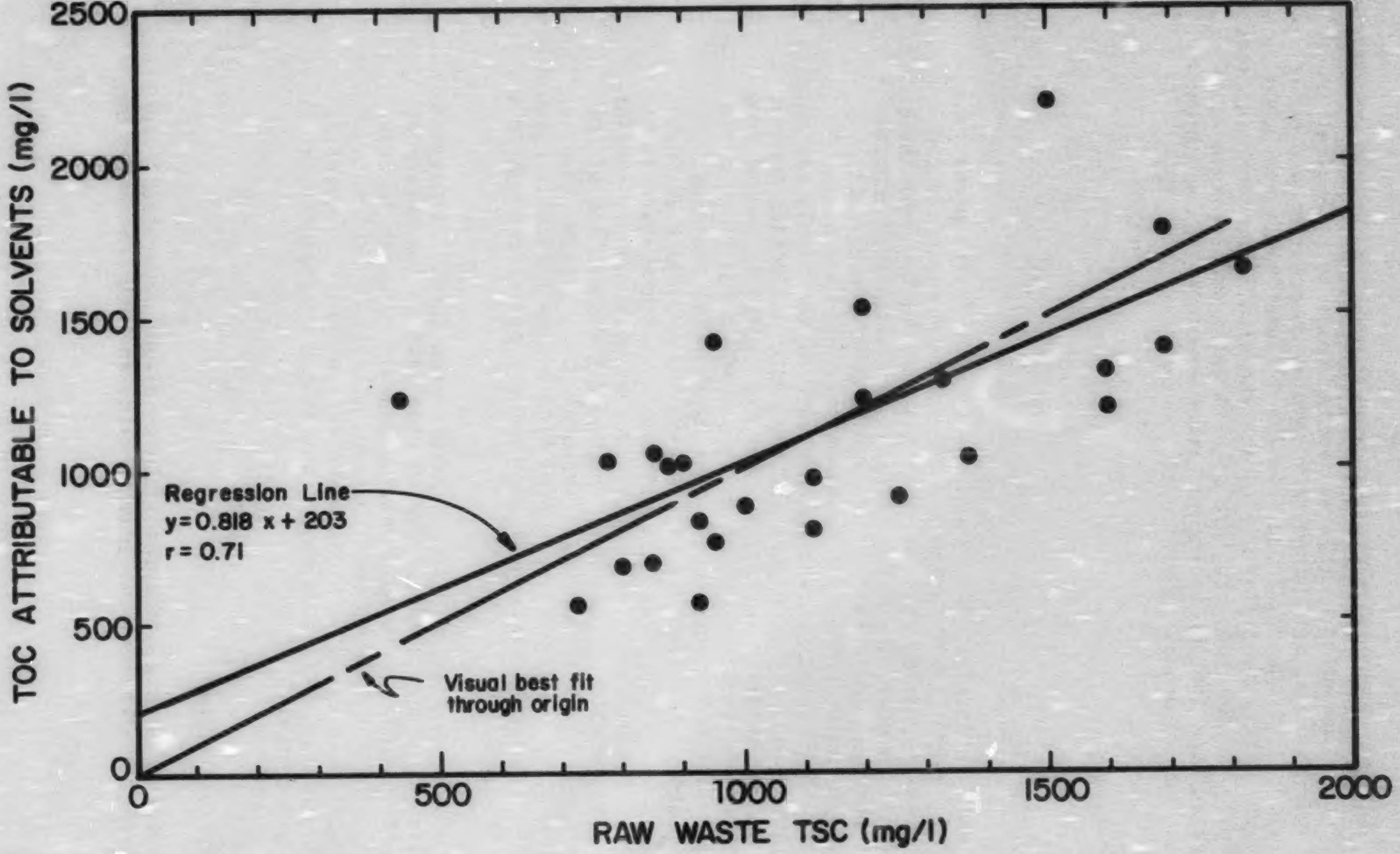


FIG. 2. CORRELATION OF MEASURED TSC TO TOC ATTRIBUTABLE TO SOLVENTS

to maintain a high level of biological treatment; therefore, nutrients must be added to this wastewater prior to biological treatment. Total solids, determined from samples composited over one-week periods, averaged 2,725 mg/l in the influent waste and 3,800 mg/l in the trickling filter effluent. The alkalinity determinations made from July 13 to July 20, are significantly greater than would be expected from the low inorganic carbon concentration and might have been due to organic constituents present in the waste.

PLANT DESIGN

Waste Collection and Transport

All process wastes from the manufacturing area in the plant flow through a central sewer system and are subsequently collected in a raw waste pumping station. All sanitary wastes from the analytical and development laboratories are discharged separately to the municipal sewer system. In addition to process wastewaters, stormwater runoff from some material storage areas is collected in the pumping station. Because of the highly variable pH of process waste discharges, the raw waste collection sump is lined with an acid-alkali resistant brick. Waste is pumped from the sump to the treatment plant using two 1,250 gpm (5 cu m/min), acid resistant pumps constructed of a high nickel stainless steel. Each pump is capable of handling the average plant waste flow of 1,000 gpm (4 cu m/min). To insure that raw waste is not discharged during power failures, a back-up power supply consisting of gas-operated generators is provided for operation of the raw waste and equalization basin influent pumps. Prior to entering the sump, all waste is screened through a mechanically-cleaned bar rack in order to remove extraneous materials discharged to the plant floor drains.

From the pumping station waste is transported through a polyester-fiberglass pipe approximately one-quarter mile to the treatment plant site. This particular construction was required in order to provide adequate resistance to acid, alkali, and solvents which might be present in high concentrations prior to equalization and neutralization. A photograph of the treatment facility is shown in Figure 3.

Preliminary Treatment

Preliminary treatment processes include solvent separation, flow measurement, neutralization, equalization, and nutrient addition.

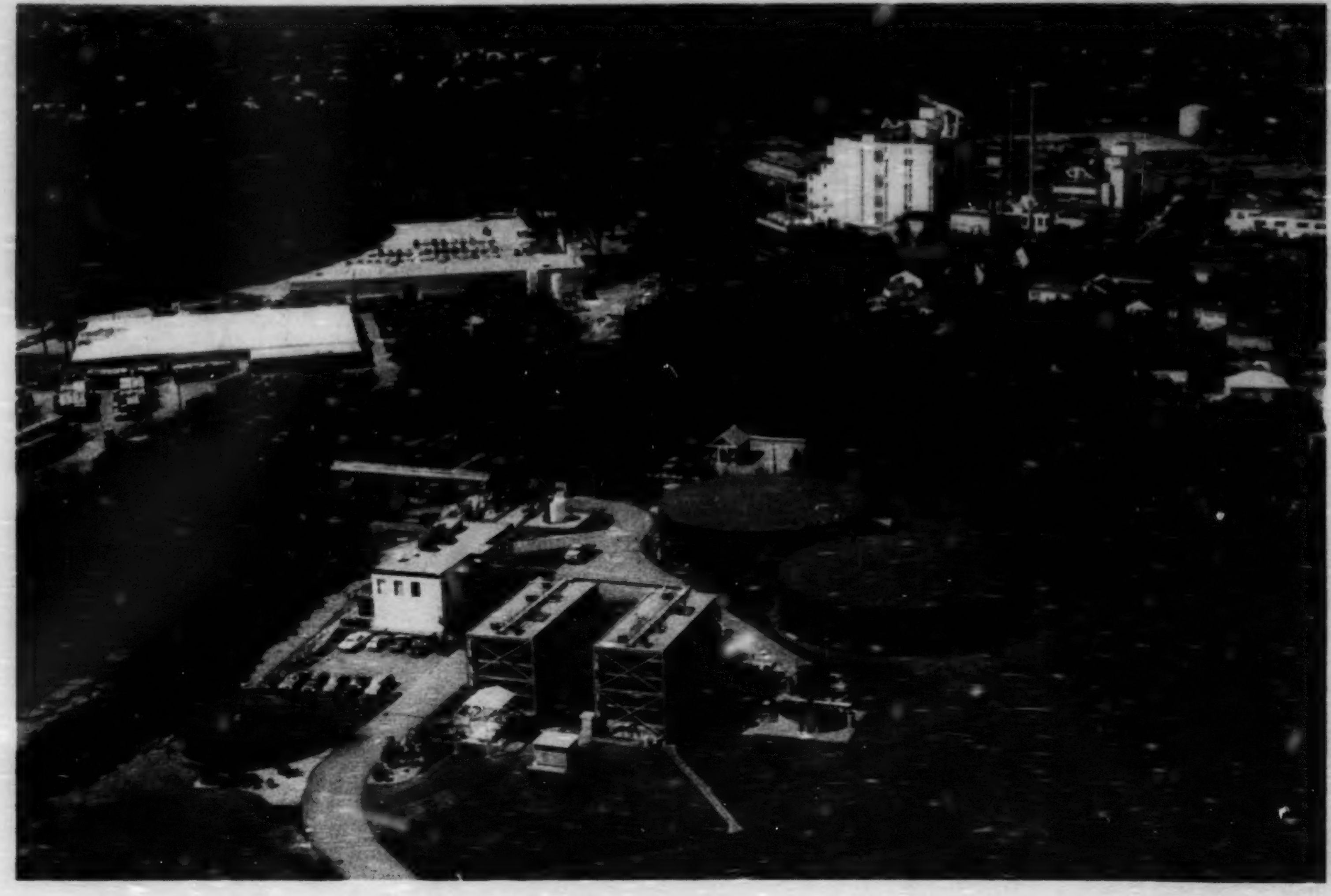


FIG. 3 .WASTEWATER TREATMENT PLANT, CIBA-GEIGY CORPORATION
CRANSTON, RHODE ISLAND

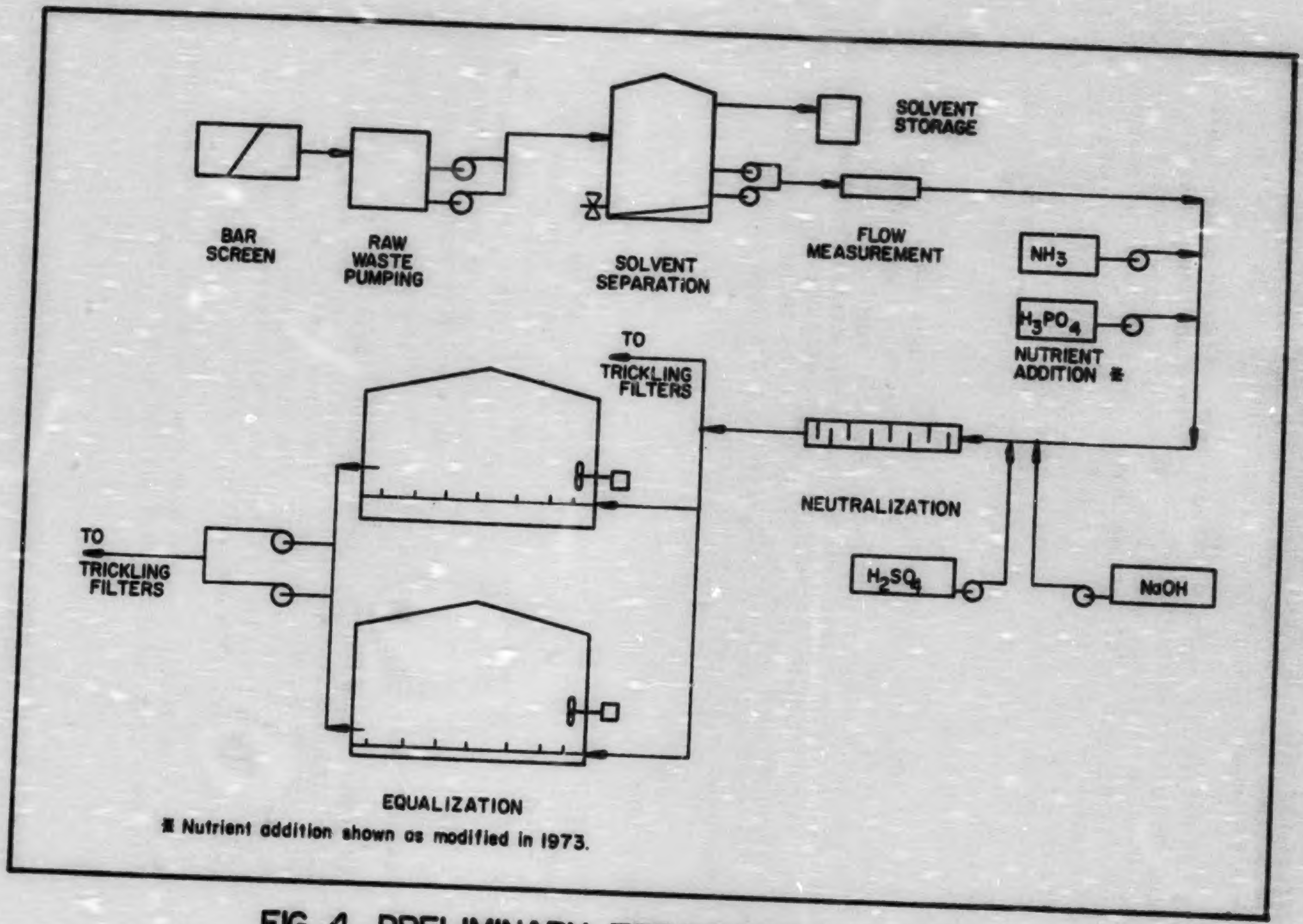
A schematic illustration of pretreatment facilities is shown in Figure 4.

Due to the possibility that large spills of solvents might occur in the plant, facilities are provided to separate immiscible solvent loads from the wastestream. This is accomplished in a 45,000 gal (180 cu m) cylindrical tank. Waste enters the tank through a tangential inlet in order to maintain most solids in suspension and yet allow separation of any immiscible solvents. In addition to serving as a solvent separator, the tank provides some equalization of pH and flow prior to the neutralization system. Immiscible solvent layers are withdrawn through a tangential outlet into a 5,000 gal (20 cu m) solvent holding tank. Collected solvents may be trucked away for ultimate disposal or transported to the manufacturing area for reclamation and reuse.

In order to provide collection of large solids which settle in the solvent separation tank, the tank bottom slopes to one side which is equipped with a valved outlet suitable for wasting sludge. Sludge is usually collected in 55-gal (200 liter) drums as necessary and taken for disposal with other concentrated wastes from the manufacturing area.

Flow Measurement and Neutralization. Waste is pumped from the solvent separation tank into the control building where the flow is metered and the waste neutralized. The waste flow is measured using a magnetic flow meter attached to the raw waste force main. Neutralization is accomplished in a one-stage, tubular reactor which was constructed by baffling an 8-ft (2.4 m) section of pipe in the raw waste force main. Probes are located at the downstream end of the reactor to monitor and control pH. Controllers are coupled to transmitters and automatic valves which control acid and caustic additions to the wastestreams. Concentrated sulfuric acid and 50 percent sodium hydroxide are used to neutralize the waste.

Equalization. Following neutralization the waste is stored in two, 1-mil gal (4,000 cu m) equalization tanks. These cylindrical steel tanks are 100 ft (30 m) in diameter with a depth of approximately 20 ft (6 m) and are located above grade. Together they provide a holding time of approximately 1.5 days for the plant design flow of 1.4 mgd (5,600 cu m/day). Each tank is lined with a phenolic material to prevent corrosion.



* Nutrient addition shown as modified in 1973.

FIG. 4 .PRELIMINARY TREATMENT FACILITIES

Flow enters the tanks through diffusers which were constructed across the bottom of each tank approximately 2 ft (0.6 m) from the bottom. Two 1,000 gpm (4 cu m/min) pumps convey the waste from the equalization tanks to the trickling filter system.

It was originally planned that the inlet diffuser system would provide sufficient mixing for equalization and would prevent solids from settling in the basin. However, a significant quantity of sludge accumulated during the first year of operation which dictated the need for mechanical mixers. Subsequently, a 40-hp (30 kw) mixer equipped with a marine-type propeller was installed in the side of each tank. Since that time no problems have been encountered due to the accumulation of solids in the tank or to poor equalization because of stratification.

Nutrient Addition. According to the original plant design, anhydrous ammonia and 75 percent phosphoric acid were added to the wastestream as it was pumped from the equalization basin to the trickling filter system. However, several problems were encountered in obtaining adequate control over nutrient addition which resulted in severe pH shocks to the biological system. Subsequently, piping modifications were made to permit nutrient addition prior to equalization.

Trickling Filtration

The trickling filtration system consists of six, 20-ft (6 m) deep trickling filters designed to operate in series. The filters were constructed in two rectangular banks with each bank containing three stages. Each stage consists of a 20 ft x 25 ft (6 m x 8 m) surface area packed with 10,000 cu ft (300 cu m) of plastic packing. The packing is supported by metal gratings in four, 5-ft (1.5 m) sections. Waste is distributed to each stage through fixed nozzle distributors.

The plastic packing is illustrated in Figure 5. The material is constructed from polyvinyl chloride and resembles a corrugated expanded metal grating. The corrugations are approximately 1 5/8 in. (4 cm) high. The packing was placed horizontally in each trickling filter tower with corrugations placed at right angles with each other in alternate layers.

Waste is collected beneath each tower in a sump having a capacity of approximately 15,000 gal (60 cu m). Since the bottom

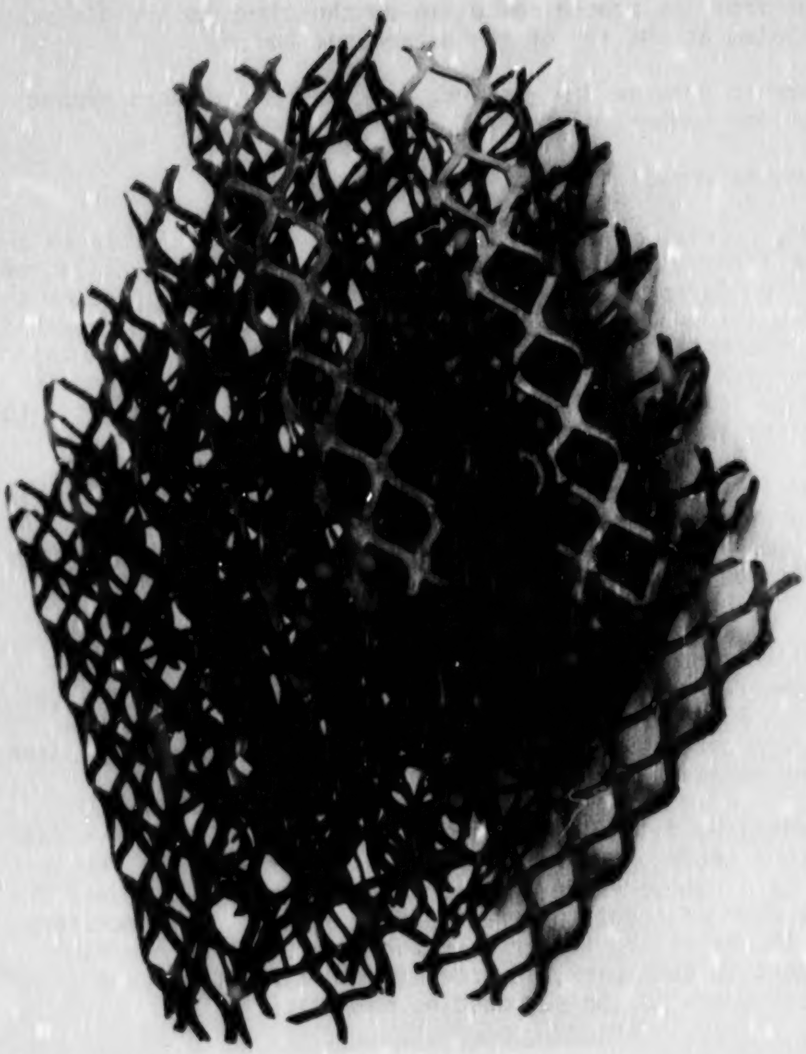


FIG. 5 . TRICKLING FILTER MEDIUM

of each sump is flat, sludge which accumulates in corners of the basin sometimes produces objectionable odors during summer months. A schematic illustration of the trickling filter system is shown in Figure 6.

Piping provided in each tower permits the following flow arrangements:

1. Under normal operation, waste is pumped from the equalization tank or from the preceding column up the riser to the distribution system at the top of the subsequent tower.

2. In order to provide for repairs, flow may be bypassed around one, or any number, of towers.

3. Flow may be recycled in any particular tower.

In addition, provisions were included to pump waste directly to the sump of the first trickling filter, thereby bypassing equalization facilities, or to recycle flow from Tower 6 to Tower 1 in order to provide flow through the system during periods when no raw waste is available. Each stage is provided with a 1,000-gpm (4 cu m/min) pump driven by a 25-hp (19 kw) motor. Pumping from each sump is controlled by level indicators in the sumps which are connected to control valves located on the outlet side of each pump.

Ventilation in the system is provided by a 10,000 cfm (28 cu m/min) blower mounted at the base of each tower which is driven by a 7.5-hp (5.6 kw) motor. Air is forced up through the trickling filter and exists through one 4 ft x 4 ft (1.2 m x 1.2 m) port located in each trickling filter bank.

Final clarification is provided in a 10-ft (3 m) deep clarifier having a diameter of 50 ft (15 m). The overflow rate for design conditions is 730 gpd/sq ft (30 cu m/day-sq m). Sludge is discharged to the city sewer and subsequently handled in the Cranston, Rhode Island municipal wastewater treatment plant.

Both the trickling filter system and the final clarifier were constructed above grade. Due to a high groundwater level in the treatment plant area, above grade construction was employed. Since the plant is located adjacent to a residential area, it was necessary to enclose the trickling filter structures, except for one air discharge port in each bank, in order to minimize the escape of biological odors to the surrounding community.

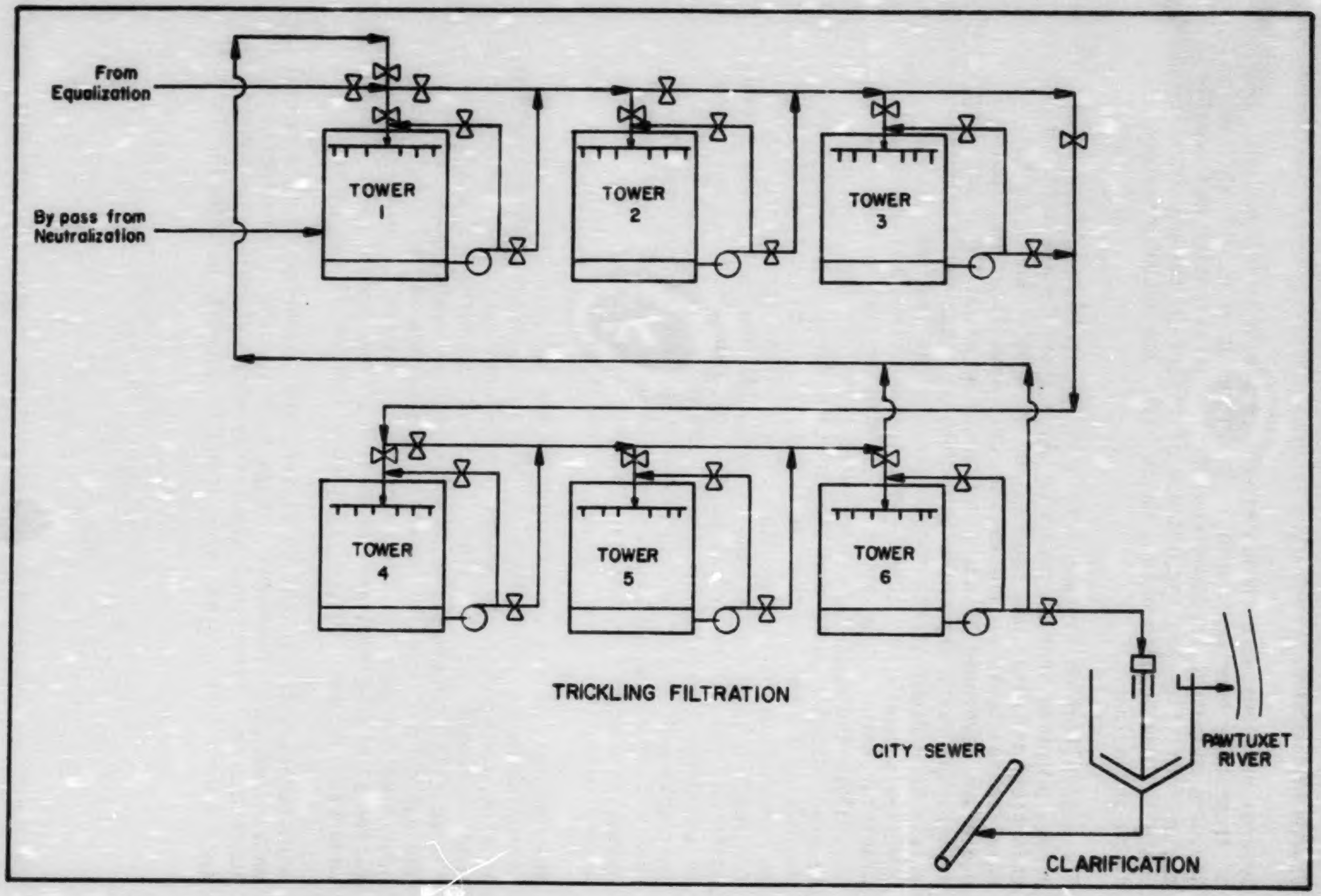


FIG. 6. SCHEMATIC ILLUSTRATION OF ORIGINAL TRICKLING FILTER DESIGN

DESIGN EVALUATION

Several problems were encountered with the original design. Two of these have been mentioned in the previous discussion: the addition of nutrients downstream from neutralization and equalization facilities and the absence of mixers in the equalization basins. Additional problems which have been encountered are discussed below.

The neutralization system, originally designed as a tubular reactor, provided poor pH control. As a short-term solution to this problem, waste was neutralized on a batch basis in the equalization tanks during most of this investigation. Using a procedure developed by the plant operator after the plant was in service, waste was introduced into one tank until approximately 80 percent of the total capacity was reached. Subsequently, a sample was taken and titrated to neutrality in the plant laboratory. The quantity of acid or caustic required to neutralize the waste was calculated from the titration and metered into the tank with the influent flow. Following a 15 min mixing period, another waste sample was taken from the tank and checked to insure that the pH was within acceptable limits for feeding to the trickling filters. As one tank of neutralized waste was fed to the trickling filters, raw waste was collected in the other equalization basin. Although it has been proven that this method can reliably be used by plant operators for controlling the wastewater pH, the capacity of the equalization system to absorb fluctuations in waste strength was sacrificed. A more suitable neutralization system is presently being designed.

The adequacy of the equalization system was not specifically evaluated during this project. While it appears that the system is probably adequate to absorb short-term fluctuations in wastewater strength and flow when the plant is operated 7 days/wk, the equalizing capacity of the system is somewhat impaired by the use of two parallel basins. Considering the difficulties in operating two parallel basins in such a manner to maximize equalization of the wastewater, the design of one larger basin would have provided more control over equalization of the waste.

Structurally, the trickling filters are supported on an external galvanized steel framework. Each 5-ft section of packing is supported on a structural steel framework extending throughout each tower. The system is designed to support an interior load of 80 lb/cu ft. Originally, the towers were wood impregnated with a decay-resistant chemical and were designed to support an interior load of 20 lb/cu ft. They suffered a complete structural failure soon after startup.

Failure of the tower system was enhanced by the lack of lateral support for the packing. Compression of the corregated-type packing when loaded with the slippery, biological slime resulted in lateral force within the columns in addition to the vertical load of the wet biota. Visual inspection of the packing, shown in Figure 5, indicates that this packing would have significantly less strength when subjected to both lateral and compression loads than compared to other commerically available plastic packing.

The capability to recycle effluent and mix it with the raw waste was not included in the original design. The advantages of effluent recirculation have been disputed in many applications. However, recirculation should be considered in treating all high-strength industrial wastes since it provides recycling of nutrients and active microorganisms to the head of the system, dilution of the influent organic concentration, and dampening of fluctuations in the influent waste load.

During the fall and winter of 1973, the system was modified to provide continuous effluent recycle. A schematic diagram showing piping modifications for recycle is shown in Figure 7. The six-stage system was modified to allow operation as two parallel units, each containing three filters in series. Flow pumped from the equalization tanks is split to Towers 1 and 4 using orifice plates and pressure-actuated control valves. Raw waste is combined with recycle flow at the top of Towers 1 and 4 and flows sequentially through three towers. Effluent from Towers 3 and 6 is split using sump level controllers (not shown in Figure 7) and an orifice plate meter, which is coupled to another control valve. A preset flow is returned to the head of the system for recycle. The balance of the flow, as sensed by the sump level monitors, is pumped to the final clarifier.

TREATMENT COSTS

A summary of treatment costs incurred for this plant are summarized in Appendix A. Table A-1 shows capital costs for the plant as estimated immediately following construction. Since the plant was built under a turnkey contract, it was not possible to obtain

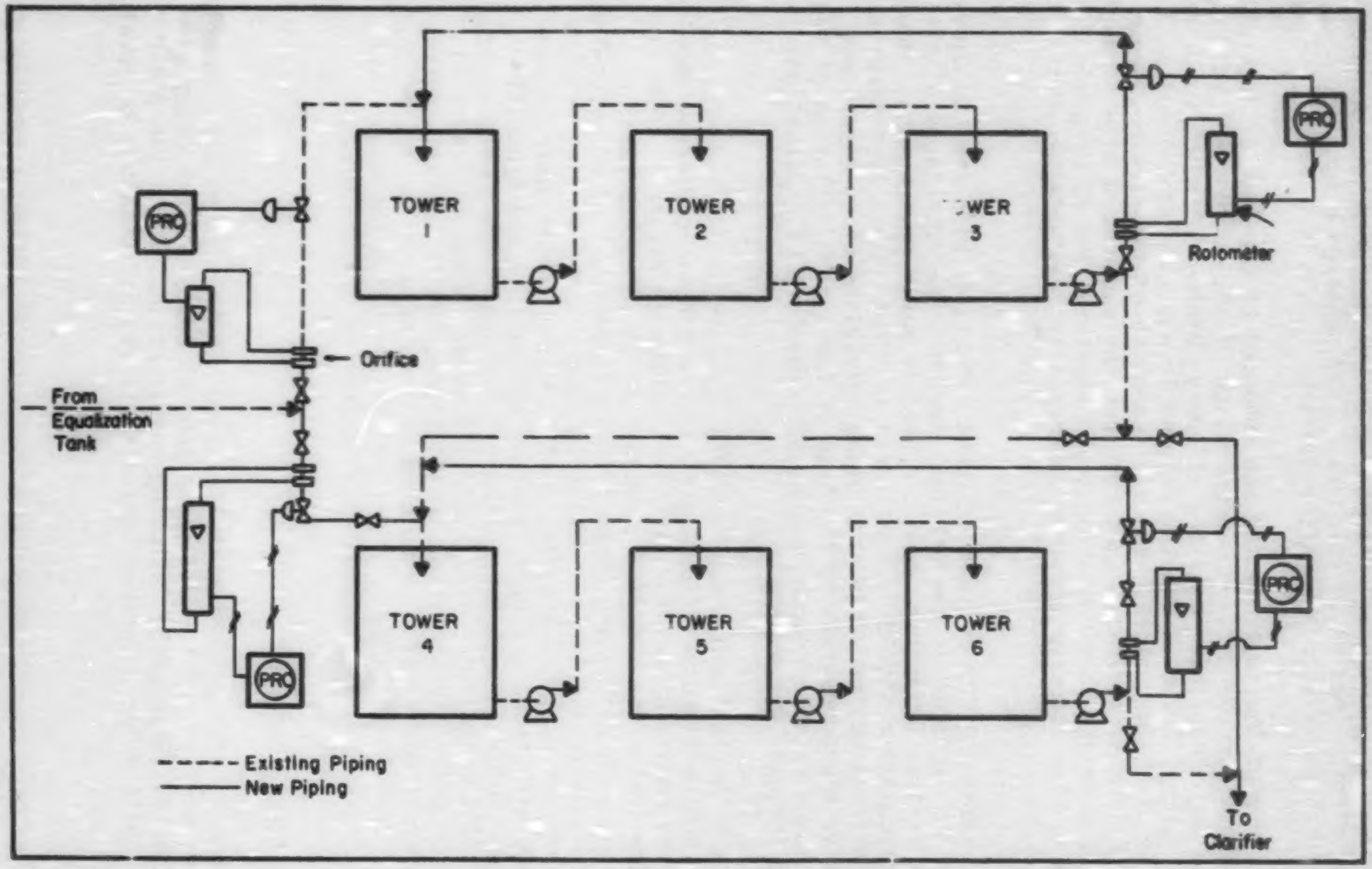


FIG. 7. TRICKLING FILTER RECYCLE SYSTEM

exact costs in all cases. In particular, no delineation of major electrical costs or piping costs (included in "Site Work") were available. It should be noted that the cost given for the trickling filters reflects the original wooden superstructure and not the steel and fiberglass structure constructed after the collapse of the original system.

The operating costs listed in Table A-2 are for 1973 which was the first year of operation following startup of the reconstructed plant. Part of the cost given for "Supervisory and Administrative Labor" included salary and other costs for the plant environmental supervisor whose duties were not completely related to treatment plant operation. While "Laboratory and Analytical" costs were somewhat high compared to other industrial plants, this level of testing will probably be required to adequately monitor a plant treating such a complex waste to insure compliance with permit limitations. Maintenance costs reflected startup problems which occurred during the year.

In order to calculate treatment costs on a volume and performance basis, it was assumed that the plant would generate 1.4 mgd of waste for 310 days/yr (6 day/wk operation) and that 11,900 lb/day of BOD would be removed for each plant operating day (equivalent to 32 percent removal from an influent BOD of 3,200 mg/l observed during this study). In order to obtain an annual cost for the capital cost of the system, an interest rate of 7 percent and a life of 10 years was assumed. Thus, the amortized cost was $263,000/yr. When combined with the annual operating cost of $327,000/yr, the total annual cost was $690,000/yr. This is equivalent to costs of $1.58/1,000 gal of waste treated and $0.18/lb BOD removed.

INSTRUMENTATION SYSTEMS

In order to determine the applicability of this system to automated control, an Ionics on-line total carbon monitoring system and a Weston and Stack dissolved oxygen system were installed in the treatment plant.

A schematic illustration of monitoring points in the system is shown in Figure 8. The total carbon analyzer and all recorders were installed in an instrument building located adjacent to Tower 3. Dissolved oxygen probes were located to monitor the contents of the effluent from each tower and the final clarifier.

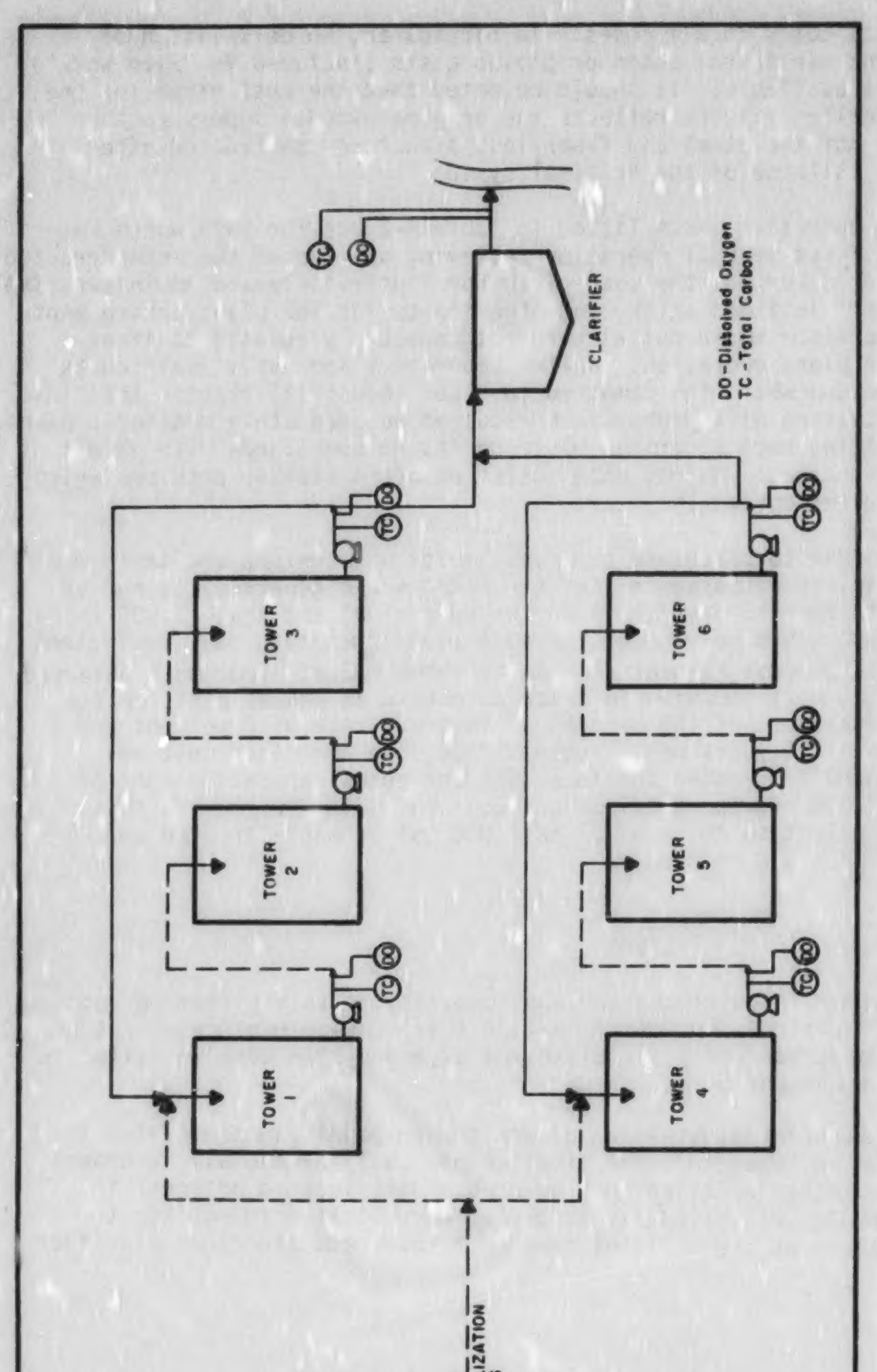


FIG.8 . WASTEWATER MONITORING SYSTEM

SECTION V

EXPERIMENTAL PROCEDURES

Due to the diversity of the tests conducted and the processes evaluated in this investigation, the experimental equipment utilized during the course of the study varied depending on the specific phase of the project. Trickling filter studies were carried out using a single-stage microtower which was small enough to be operated indoors. A two-stage pilot trickling filter was constructed on-site at the CIBA-GEIGY plant for use in other parts of the study. Activated sludge investigations using air aeration were performed in 40-gal (150 liter) pilot units which were later sealed for use in the oxygen aeration investigation. Activated carbon tests were made using 1-in. (2.5 cm) and 4-in. (10 cm) column units. In this chapter the equipment and operating procedures, analytical methods, and data correlation procedures employed during the study are described.

CONDUCT OF THE INVESTIGATION

A schematic drawing of the present treatment facilities showing points at which feed was taken for experimental units is shown in Figure 9. Activated sludge investigations were performed using raw equalized waste, unclarified trickling filter effluent, and clarified trickling filter effluent in order to determine any differences in the response of these wastes to activated sludge treatment. Granular activated carbon was tested using both the clarified trickling filter effluent and the activated sludge effluent. All trickling filter and air stripping investigations were performed using raw equalized waste.

Because the principal objective of the pilot trickling filter and microtower studies was to determine the effect of various operating modifications on the full-scale treatment system, all tests were conducted using equalized raw waste. A continuous feed of raw waste from the equalization tanks was pumped to the pilot trickling system, while feed for the microtower was taken from the equalized raw waste sampler at 24-hr intervals. Therefore, the feed to the microtower was subjected to fewer concentration fluctuations than was the pilot trickling filter system.

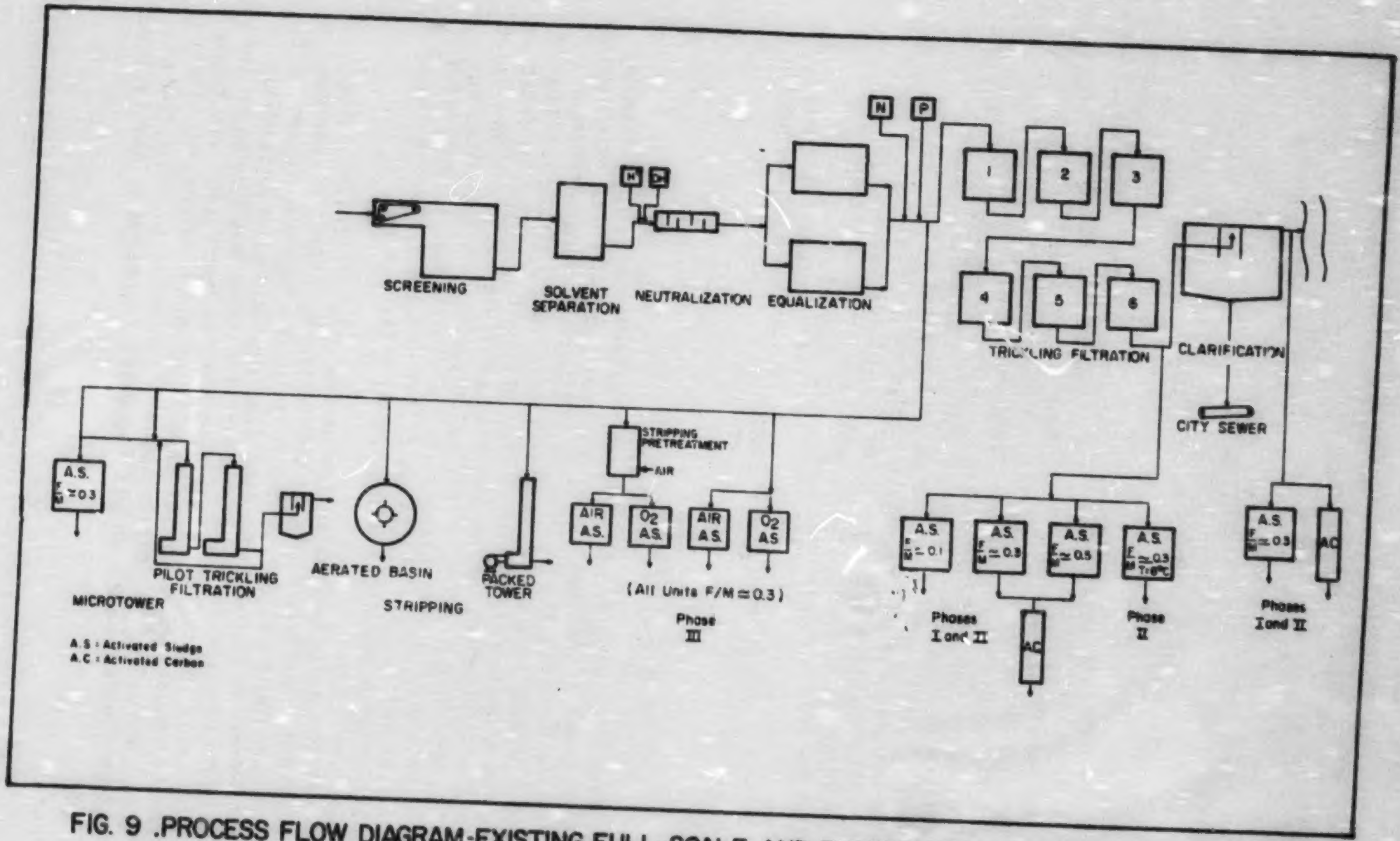


FIG. 9 .PROCESS FLOW DIAGRAM-EXISTING FULL-SCALE AND EXPERIMENTAL TREATMENT SEQUENCES

The testing program with the microtower included the use of: 1) raw waste, 2) raw waste diluted with tap water and, 3) effluent recycle. All tests in the pilot trickling filter used raw waste as feed to the system and effluent recycle.

Activated sludge investigations were divided into three phases. Phase I was conducted during January and February, 1973 and had the objective of determining the feasibility of activated sludge treatment for this waste. During Phase I investigations, tests were conducted using raw wastewater, unclarified trickling filter effluent and clarified trickling filter effluent. Phase II investigations were conducted during the months of March, April and May and had the objective of determining the response of the activated sludge system to cold temperatures (during March and April) and of determining the longer term response of activated sludge treatment to this waste. Based on favorable results from Phase I, unclarified trickling filter effluent was used for the Phase II investigation. The objective of Phase III activated sludge investigations was to compare air and oxygen aeration for the treatment of this wastewater. During this portion of the investigation, air and oxygen activated sludge units were fed unstripped and stripped equalized waste.

During the Phase I activated sludge investigation, units were operated at food-to-microorganism (F/M) ratios of 0.1, 0.3, and 0.5 g BOD/day-g MLVSS. In the remaining portions of the study, the loading of all units was maintained relatively constant at an F/M of approximately 0.3.

Activated carbon tests were conducted to determine the feasibility of adsorption for removing color or gross organics from both the trickling filter and activated sludge effluents. Isotherm tests were initially made using eight different varieties of carbon. Based upon the results of these tests, three carbons were chosen for further study in column contacting systems.

Air stripping investigations were conducted to determine the removals of organic constituents which could be achieved in an aerated basin or a packed tower. A 13,500-gal (51,000 liter) basin was constructed to simulate an aerated equalization basin, while one pilot trickling filter was used to simulate air stripping in a packed tower. Raw equalized waste was used as a feed for both experimental investigations. By conducting Phase III activated sludge investigations using both stripped and unstripped waste, effects of stripping on the response of the waste to activated sludge treatment were determined.

DATA CORRELATION

Data collected in this investigation were generally correlated using established mathematical models. In trickling filter investigations, the model which most closely represented experimental findings was a first-order model used by Eckenfelder (6, 16). Activated sludge data were correlated using the models which most closely conformed to those proposed by Eckenfelder (17, 18, 19) and Pearson (20). The effects of temperature were corrected using the Streeter-Phelps model (21).

EXPERIMENTAL EQUIPMENT

Pilot Trickling Filter

A schematic diagram of the pilot trickling filter used in this study is shown in Figure 10. This system consisted of 2 ft x 2 ft (60 cm x 60 cm) towers containing 20 ft (6 m) of the Monsanto Biodize trickling filter medium. The plastic medium was supported in four, 5-ft (1.5 m) sections on expanded metal grating. Each tower was supported above a reservoir with a capacity of approximately 500 gal (2,000 liters) and was equipped with a forced draft ventilation system adjusted to deliver 100 cfm (2.8 cu m/min) of air through the tower packing. Spray nozzles were placed above the media in each tower at a distance to provide proper wetting of the media.

Provision was made for piping raw equalized waste to a holding reservoir which was used as the influent for the system. Raw waste feed to the first column was controlled using a Fisher and Porter flow controller. A level controller on the reservoir insured that an adequate supply of feed was present at all times. Pumping was provided for the influent to the first tower, from the first to the second tower, from the second tower to the clarifier, and for recycle of effluent from the second tower to the top of the first tower. It was possible to adjust all pumps to deliver flows ranging from 1 to 10 gpm (4 to 40 liters/min). Level controllers were also installed in the reservoir beneath each tower.

The system was also provided with a 5-ft (1.5 m) clarifier. Sludge wasting from the clarifier was controlled by automatic timers.

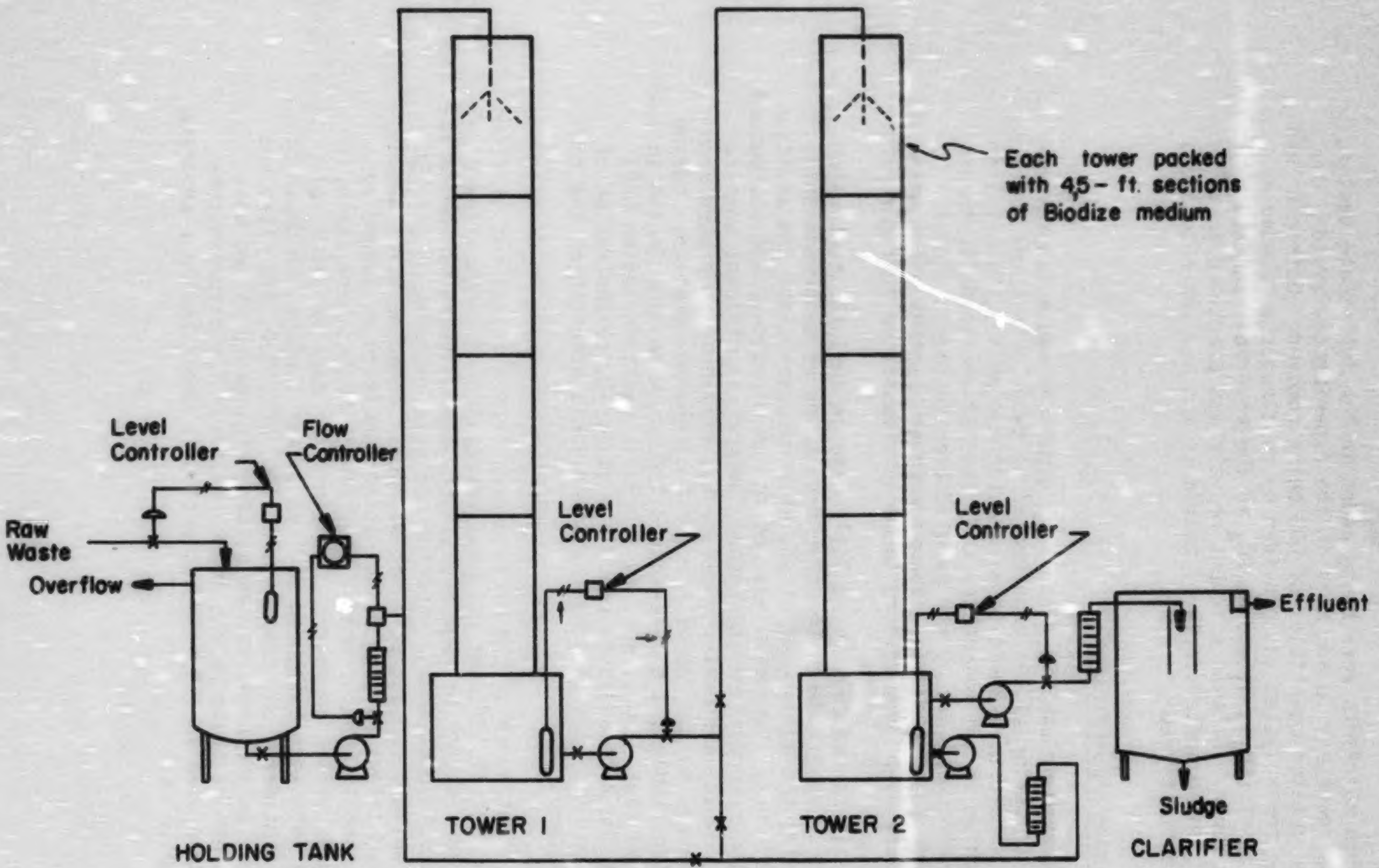


FIG.10. SCHEMATIC ILLUSTRATION OF PILOT TRICKLING FILTER
(m= ft×0.3)

Although nutrients were being added in the full-scale system, it was necessary to add more in the experimental systems due to the greater quantities of BOD being removed. Nutrient feed to the tower system was provided by continuously pumping a mixture of $(NH_4)_2SO_4$ and Na_3PO_4 to the reservoir beneath the first tower. Because effluent recycle was practiced in all tests, an adequate supply of nutrients was available to biota in the first tower.

Microtower Apparatus

The microtower used in these investigations was a laboratory-scale trickling filter constructed in two, 10-ft (3 m) sections. which simulated a 20-ft (6 m) plastic medium trickling filter. The medium in the microtower consisted of inclined PVC sheets which were designed by the manufacturer to simulate B. F. Goodrich Koroseal media. The microtower assembly along with auxiliary pumps and holding tanks is illustrated in Figure 11.

Equalized raw waste was fed daily into an influent tank which was continuously mixed to prevent settling of suspended solids. Following the adjustment of pH, when necessary, and the addition of nutrients, this waste was fed to the microtower and was pumped from the reservoir on the second tower to an effluent holding tank. During effluent recycle tests, effluent was pumped from this barrel to the top of the first microtower section. During dilution studies, the necessary dilutions were made in the influent feed tank. Supplementary inorganic salts including NaCl and Na_2SO_4 were added to produce a mixed TDS concentration of 2,000 mg/l to simulate the salts and ionic strength of the raw wastewater.

Daily maintenance of the microtower included adjustment of flows, measurement of effluent volumes, correction of uneven flow distribution pattern through the filter, and cleaning of the influent and effluent drums. Routine analyses included pH and TSC daily on influent and effluent samples. When the tower was acclimated to a particular operating condition as judged by consistent TSC measurements, additional samples were taken. Under stabilized operating conditions for each run, the following data were collected at five points through the depth of the microtower: pH, temperature, TSC, COD, BOD, and dissolved oxygen. In addition, oxygen uptake rates of the slime were determined at these points in the filter. The oxygen uptake procedure used for this investigation is described in a separate section below.

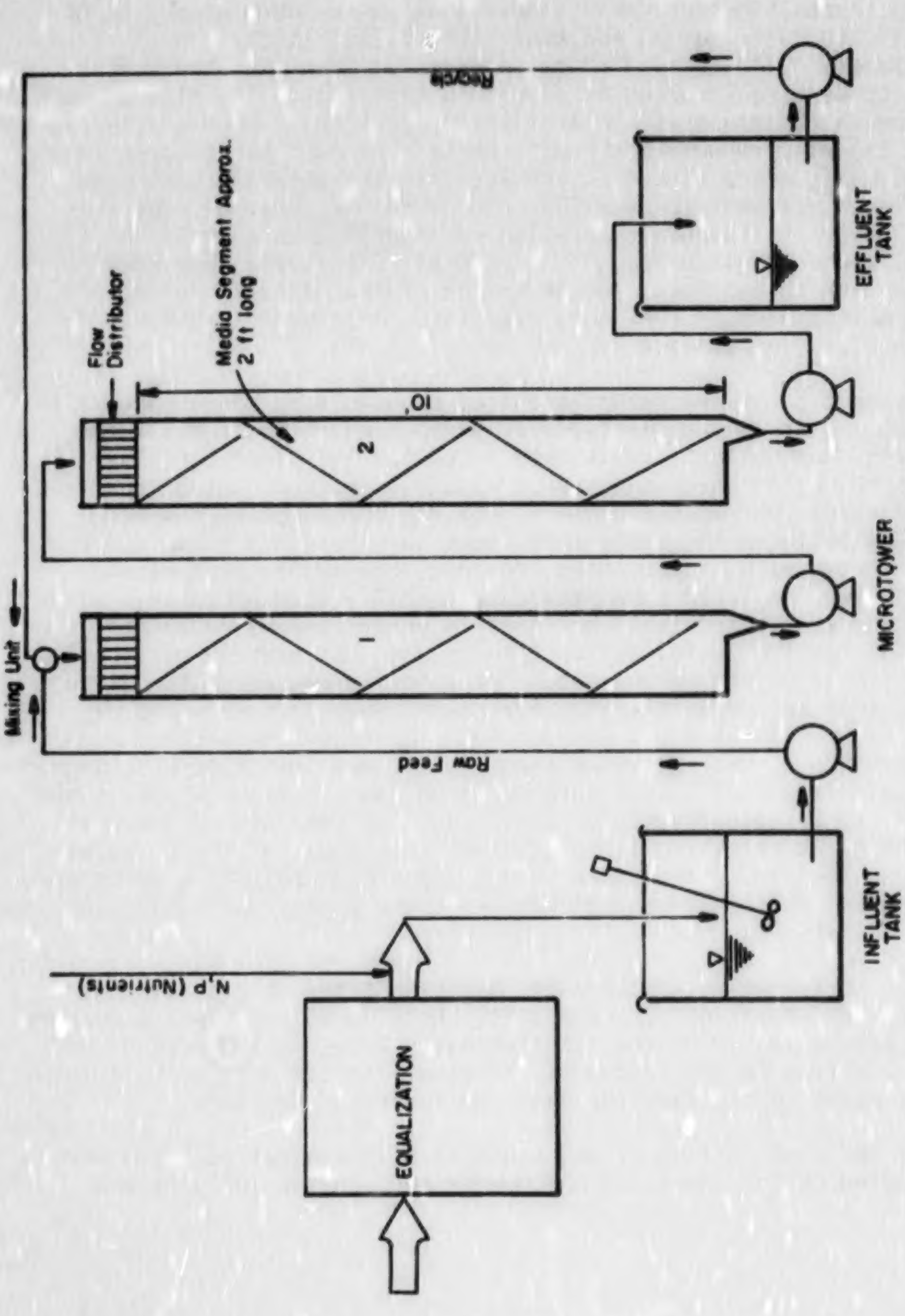


FIG.11. SCHEMATIC OF THE MICROTOWER ARRANGEMENT
(m= ft x 0.3)

Activated Sludge Units

Continuous-flow activated sludge units were fabricated from 55-gal (200 liter) drums and had a 40-gal (151 liter) capacity. A schematic illustration of these units is shown in Figure 12. Units were constructed by welding a baffle into the side of each drum which served as a clarification section. Three-quarter inch (2 cm) nipples welded through the side of each barrel permitted the maintenance of a constant level in the aeration basin and allowed effluent to overflow into containers provided for this purpose. Diffused air aeration was supplied through 12-in. (30 cm) porous stone diffusers. These diffusers were suspended in the tank to provide adequate mixing of the mixed liquor solids, to maintain desired oxygen levels and to prevent turbulence in the clarifier section.

The feed to the units was provided through a variable-speed tube pump. Effluent, collected from various points in the treatment plant, was collected daily and added to feed drums. Effluent was collected in containers provided for each unit and the effluent volume was measured daily. Nitrogen and phosphorus were added daily to the waste feed tanks in order to provide a BOD:N:P ratio of 100:5:1. When required, the pH of the influent waste was adjusted to approximately pH 7. When necessary, mixed liquor was wasted from the units in order to provide an operating MLVSS level of 2,500 mg/l. After sludge was wasted, effluent collected during the previous 24-hr period was added to make up the lost volume in the aeration basin.

Operation of the activated sludge units was initiated by using an equal mixture of sludge obtained from the Cranston, Rhode Island municipal activated sludge plant and clarifier sludge from the CIBA-GEIGY trickling filter system. Operation of the units was begun the first of December, 1972. Approximately five weeks were required for complete acclimation of the biological solids to this waste.

Flow to the units was adjusted daily in order to maintain the desired F/M ratios for each unit. Upon occasion it was necessary to adjust the pH of the aeration basin by adding 50 percent NaOH. In addition it was frequently necessary to add a silicone defoaming agent to the aeration basin to suppress frothing.

The influent, effluent, and contents of the aeration basin were sampled daily. Analyses which were made on the influent and

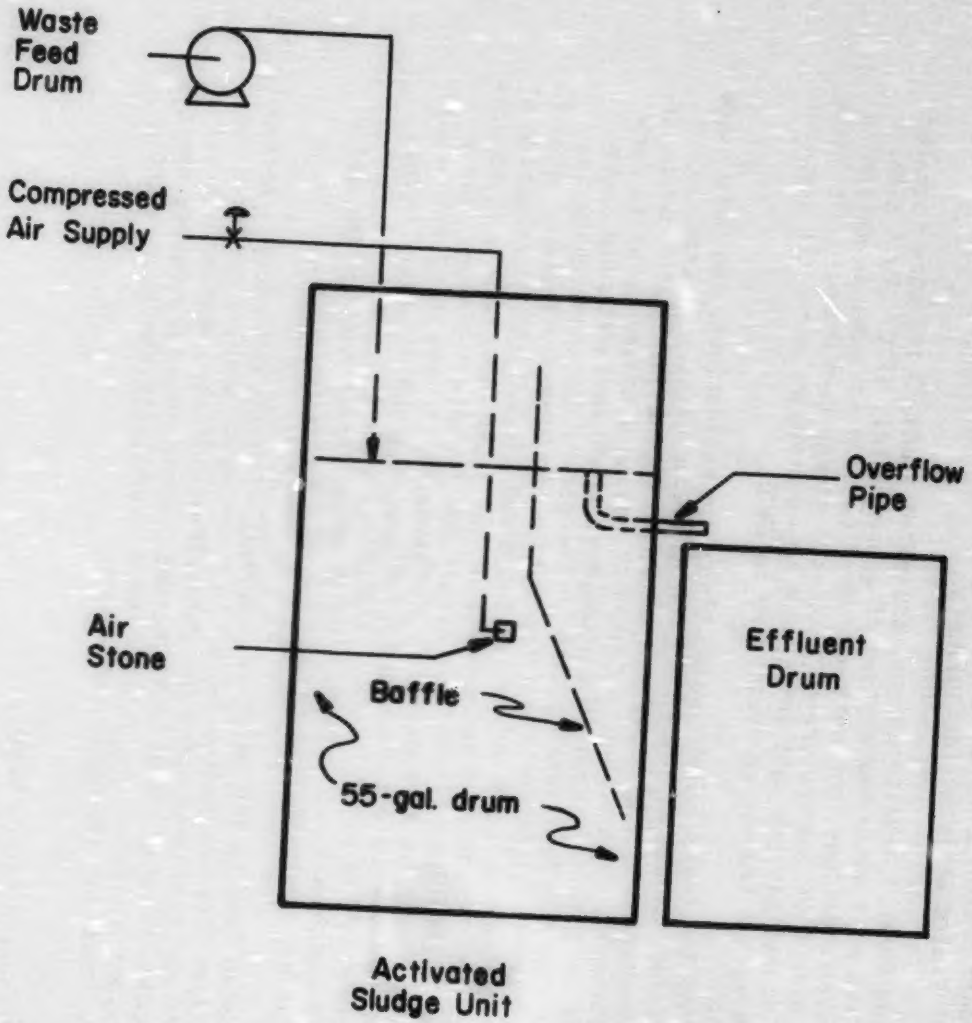


FIG. 12. SCHEMATIC ILLUSTRATION OF ACTIVATED SLUDGE UNITS
(Liters = gal × 3.8)

effluent samples daily included pH, TSC, and suspended and volatile suspended solids. These samples were analyzed three times per week for total and soluble BOD and COD.

Contents of the aeration basin were analyzed daily for total and volatile mixed liquor suspended solids and pH. These samples were analyzed four times per week for oxygen uptake rate, zone settling velocity, and sludge volume index. Weekly samples were taken for total kjeldahl nitrogen and total phosphorus measurements. Occasionally, samples were analyzed for solvents.

During Phase I investigations, activated sludge units were operated indoors at a waste temperature of approximately 17°C. During Phase II investigations, one unit was moved outside in order to determine the effect of temperature on activated sludge performance. In Phase III investigations, all units were moved outdoors where the temperature varied from approximately 20 to 24°C.

During the Phase III activated sludge investigations, two units were modified for use as oxygen activated sludge units. A schematic illustration of these units is shown in Figure 13. Modification of these units essentially consisted of adding a sealed cover to each unit so that an oxygen-rich atmosphere could be maintained above the liquid. Additional piping in the covers for these units was provided to allow for the introduction of oxygen, withdrawal of off-gases, recirculation of gas to the mixed liquor, and introduction of feed to the unit. A gas compressor was piped into each unit to provide a constant diffused air supply for the mixed liquor. These compressors had a capacity of approximately 1 scfm (0.03 cu m/min) and were adequate to maintain proper mixing and dissolved oxygen levels in the mixed liquor. Additional piping was added to allow the withdrawal and introduction of mixed liquor samples for daily testing.

Oxygen feed to the units was supplied from gas cylinders. Influent oxygen flow was controlled using a stainless steel metering valve and was monitored using gas rotometers. Off-gas measurement was accomplished using wet test meters. Operational procedures for the oxygen activated sludge units were essentially the same as those followed for air activated sludge.

Activated Carbon Tests

Carbon Isotherms. Batch adsorption tests were conducted by contacting waste samples and granular activated carbon in a wrist-action

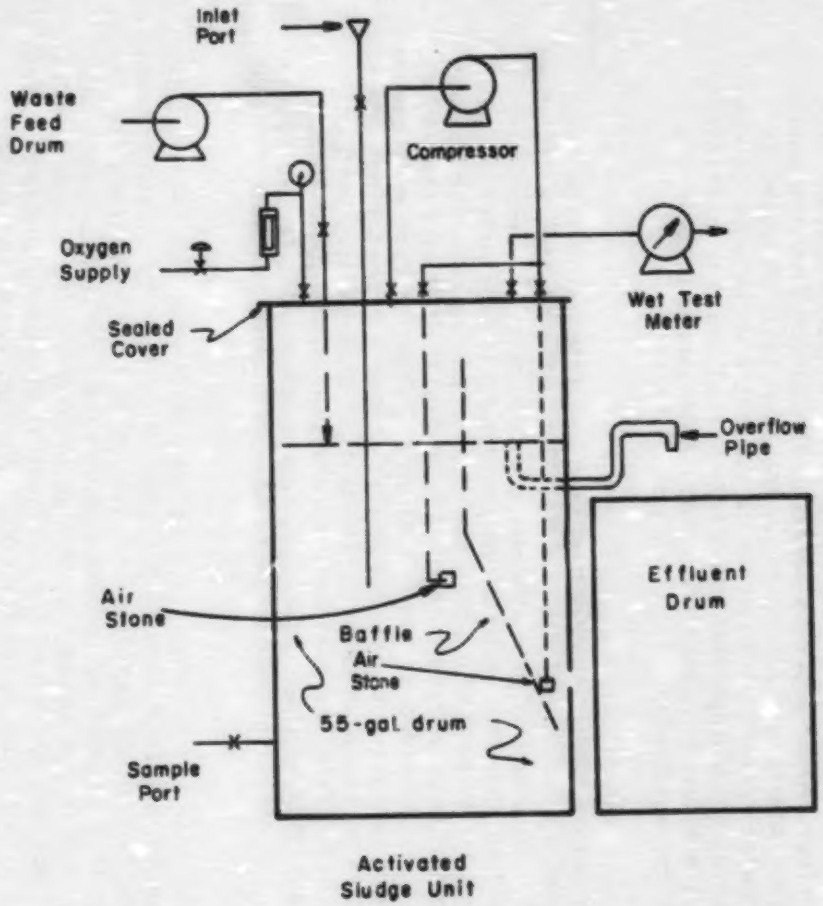


FIG.13. SCHEMATIC ILLUSTRATION OF OXYGEN ACTIVATED SLUDGE UNITS
(Liters = gal × 3.8)

laboratory shaker for a 24-hr period. Although powdered carbon is frequently used for isotherm tests, results using the granular carbon are more representative since waste constituents must diffuse through pores in the carbon particles before being adsorbed onto the carbon. Granular carbon was sieved to obtain the desired range of particle sizes, soaked in distilled water for 24 hr, oven dried at 103°C for 24 hr, and cooled in a dessicator. Varying amounts of carbon were added to each test flask, and then contacted for 24 hr with 200 ml of waste which had been previously filtered through Reeve Angel Grade 802 filter paper. Usually carbon was added to obtain COD to carbon ratios ranging from 0.05 to 5.0 g/g.

After 24 hr the waste-carbon mixture was filtered through GF/C glass fiber filter paper to separate the carbon from the waste. A control sample was subjected to the same test procedure as each carbon sample. Initial tests conducted using varying contact times determined that the 24-hr period would be sufficient for all practical purposes to obtain equilibrium between the carbon and wastewater.

Column Units. One carbon column unit was fabricated from 1-in. (2.54 cm) Pyrex glass tubes and is illustrated in Figure 14. Feed to these columns was provided using a tube pump. Columns were operated upflow with each column containing approximately 3 cm of glass beads supporting 60 cm of activated carbon. The columns were operated at a flow rate of 2.5 gpm/sq ft (0.1 cu m/min-sq m). Clamps were provided at the top of each column so that each carbon bed could be backwashed individually. Solids accumulation made it necessary to backwash each column daily. Grab samples were collected daily from the influent and effluent from each column. Daily analyses were made for TSC, suspended solids, turbidity, and color. In addition, samples were analyzed twice per week for BOD and COD.

A set of four, 4-in. plexiglass columns were used for testing adsorption of the trickling filter effluent. Feed for these columns was provided by constantly pumping clarifier effluent from the full-scale treatment system into the control building. Waste was pumped through the columns at a rate of 5 gpm/sq ft. Grab samples of the influent waste and the effluent from each column were taken periodically and analyzed for COD, BOD, TSC, color and suspended solids. Columns were backwashed as necessary to prevent excessive headloss.

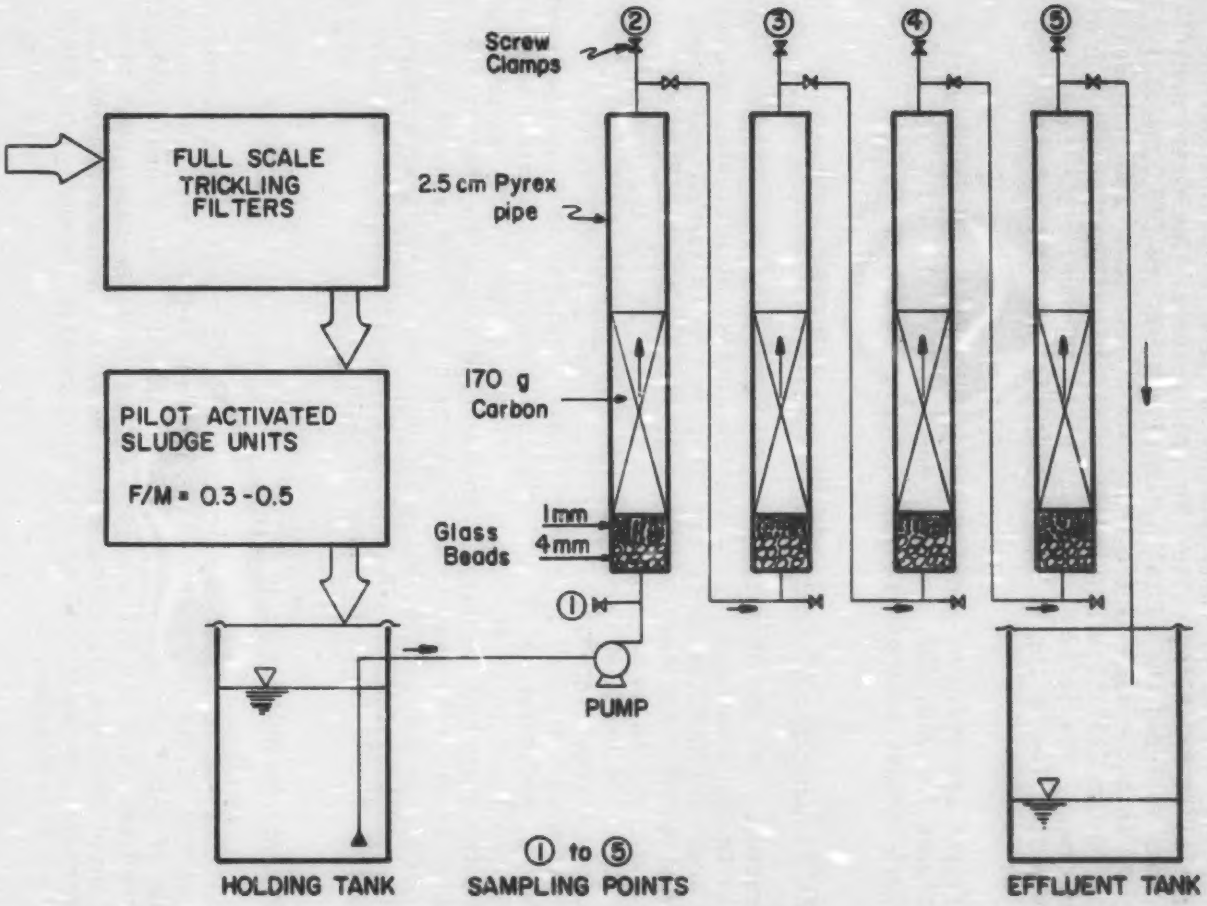


FIG. 14. SCHEMATIC ILLUSTRATION OF ACTIVATED CARBON COLUMNS

Packed Stripping Tower

The removal of organic constituents which could be achieved by air stripping in a packed tower was investigated by converting the first pilot trickling filter for this purpose. Necessary modifications included the installation of a 2,500 cfm (70 cu m/min) blower into the side of the reservoir of the tower and the addition of a 40-gpm (151 liter/min) pump to supply influent to the top of the tower. The influent pump was piped to receive the equalized raw waste flow in the influent reservoir for the pilot trickling filter system.

Influent and effluent samples were collected during each run and analyzed for BOD, COD, and suspended solids. In addition, occasional samples were analyzed for solvent concentrations.

Aerated Equalization Basin

Because of the possibility of adding additional equalization capacity to the treatment plant, the feasibility of achieving removal of organic constituents by air stripping in an aerated basin was investigated. For this purpose a 24-ft (7.3 m) diameter, 5-ft (1.5 m) deep swimming pool was constructed at the treatment facility. The basin was designed for a water depth of 4 ft (1.3 m) and had a resulting capacity of 13,500 gal (51,000 liters). Aeration was provided by a 2-hp (1.5 kw) aerator (Mixing Equipment Company, Rochester, N. Y.). This corresponded to a power level of 150 hp/mil gal (0.03 kw/cu m). Influent to the basin was provided from the equalization basin. An inverted U-tube located at the outlet of the basin was used to maintain a liquid level of 4 ft (1.2 m). Samples of the basin influent and effluent were taken daily and analyzed for BOD, COD, and suspended solids. An illustration of this basin is shown in Figure 15.

SAMPLING AND ANALYTICAL METHODS

Analytical methods used during the course of this investigation conformed to those prescribed in Standard Methods for the Examination of Water and Wastewater (13th edition). The graduate dilution method was used for BOD analyses except in a few cases when time dictated that the pipetting method be used. Soluble BOD and COD determinations were made by filtering samples through glass fiber (GF/C) filter paper.

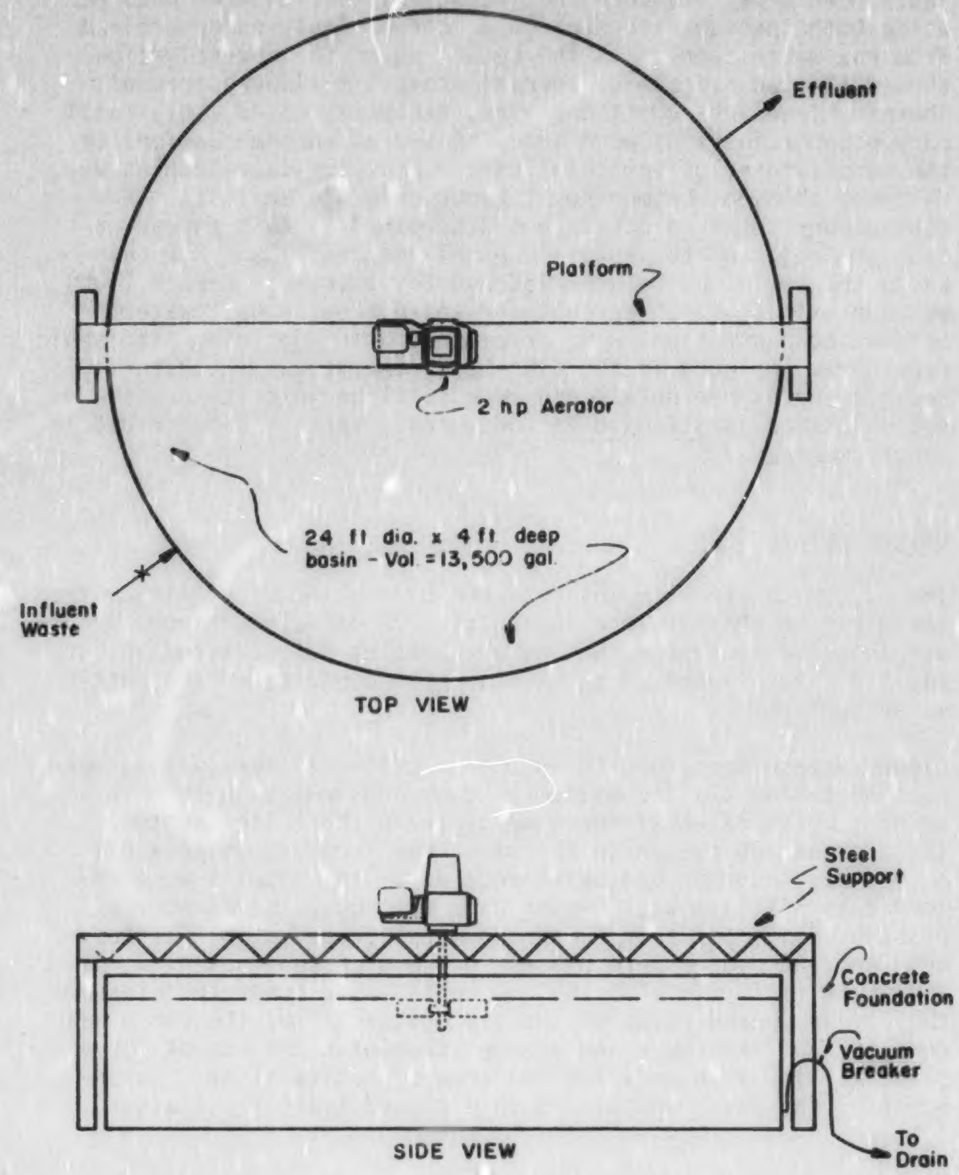


FIG. 15. AERATED EQUALIZATION AND STRIPPING BASIN
(m = ft × 0.3; kw = hp × 0.77; cu m = gal × 0.004)

Total carbon analyses were initially made using an Ionics total carbon analyzer and later using a Beckman Model 915 analyzer. Tests were made to insure that identical results were obtained using both instruments. Inorganic carbon analyses determined from raw waste samples at the beginning of the investigation showed that an average of approximately 5 mg/l were present. Therefore, in subsequent analyses, except as noted, only total carbon determinations were made. Based on recommendations of the manufacturer of the total carbon analyzer, all samples were filtered through Whatman No. 2 paper prior to analysis. For consistency total carbon values determined in this investigation are referred to as total soluble carbon, TSC. For this waste the variation between TSC and total organic carbon (TOC) was approximately 0.5 percent for the untreated wastewater. Solvent concentrations were determined using gas chromatographic procedures employed in the CIBA-GEIGY analytical laboratory. Because the oxygen uptake and zone settling velocity tests are not described in Standard Methods, these tests are described in detail separately.

Oxygen Uptake Test

The oxygen uptake rate was measured by placing mixed liquor from the aeration chamber in a BOD bottle. A dissolved oxygen probe was inserted, insuring that no air bubbles were trapped inside, and the mixer turned on to maintain the contents of the bottle in suspension.

Oxygen uptake measurements in the trickling filter systems were made by taking plastic medium patches and immersing them in waste samples taken at the same depth in the filter system. The decrease of oxygen in the waste was recorded using a dissolved oxygen probe and chart recorder. The results were recorded as milligrams of oxygen used per square centimeter of plastic medium per hour. For the microtower system, patches cut from the medium were placed in the microtower at 4, 8, 14, and 16-ft (1.2, 2.4, 4.3, 4.9 m) depths. In order to determine the oxygen uptake rates of the biological slime, the patch was removed from the tower and scraped free of slime except for a 3-in. x 3-in. (7.6 cm x 7.6 cm) area of active slime. Subsequently, the patch was placed in a closed 1-liter container filled with waste and the oxygen depletion recorded with time. Patches were located in the microtower so that they could be replaced after each determination.

To make oxygen uptake determinations in the full-scale trickling filter system, a 5-in. x 5 3/4-in. (13 cm x 15 cm) piece of plastic medium was cut from a larger section located at mid-depth in each tower. Care was taken to obtain sample patches located far enough into the tower to receive a representative portion of waste flow.

Zone Settling Velocity

The zone settling velocity test (ZSV) expresses the rate of interface settling of particular sludge in ft/hr (m/hr). The ZSV test is conducted by filling a 1,000-ml cylinder with mixed liquor suspended solids from the aeration chamber. The contents were slowly stirred with a mechanical device at a speed of 4-6 revolutions/hr. This stirring was used to break up the bridging of the sludge in the small-diameter cylinders. As the sludge settled, the interface was recorded at selected time intervals. The interface level was plotted against settling time and the zone settling velocity calculated from the slope of the straight line portion of the curve.

SECTION VI

SMALL-SCALE TRICKLING FILTER INVESTIGATIONS

Investigations in small-scale trickling filter systems were undertaken to examine and verify BOD removal mechanisms which were occurring in the full-scale filter system, and to determine the effects of recycle on filter performance. Tests were made both with a single-stage microtower operated in the laboratory and a two-stage plastic medium pilot trickling filter. Investigations were conducted using the microtowers in order to gain increased operating flexibility compared to the pilot system.

MICROTOWER TESTS

Equalized raw wastewater, collected continuously from the influent to the full-scale system, was fed daily to the microtower. Investigations in the microtower system consisted of three parts: 1) tests using a raw waste feed, 2) tests using a raw waste feed diluted with 0.5, 1, or 2 parts of tap water, and 3) tests in which filter effluent was recycled along with raw waste. Removable sections of plastic packing were placed into the microtower at several depths in order that oxygen uptake determinations could be easily made.

Effect of Variations in Organic Loading

A range of raw waste flows varying from 0.66 to 3.10 gpm/sq ft (0.03 to 0.13 cu m/min-sq m) was applied to the filter in order to study its response to varying concentrations and organic loadings. The influent COD concentration for these tests varied between approximately 4,000 and 6,600 mg/l. The corresponding organic loading varied between approximately 1,000 and 13,000 lb COD/day-1,000 cu ft (16 and 200 kg COD/day-cu m). In dilution studies, raw waste feed was diluted with 0.5 to 2.0 parts of tap water. Applied organic loadings in these tests varied from approximately 1,000 to 6,000 lb COD/day-1,000 cu ft (16 to 96 kg COD/day-cu m) with applied COD concentrations ranging from approximately 750 to 2,400 mg/l COD. Hydraulic application rates for the dilution tests were either 1.4 or 2.1 gpm/sq ft (0.06 or 0.09 cu m/min-sq m). Thus, in these tests the applied COD concentration for corresponding loading rates (lb COD applied/day-1,000 cu ft or kg COD applied/day-cu m) was significantly lower compared to tests in which undiluted raw waste was applied to the filter. The results of these tests, shown in Figure 16, reveal that significantly greater percentage COD removals were obtained at corresponding organic loadings when the undiluted raw waste was applied to the filter. Differences in performance might be due to either differences in hydraulic loading or to the concentration differences. It is felt that concentration differences largely explain the difference in performance due to reduced removals by air stripping at the lower applied concentrations.

In the effluent recycle studies, hydraulic loadings applied to the filter ranged from 1.9 to 2.8 gpm/sq ft (0.08 to 0.11 cu m/min-sq m). The COD removal efficiency for these tests, shown as a function of applied organic loading, is illustrated in Figure 17. In this case, the removal was calculated based on the organic concentration of the raw wastestream. It can be seen that organic removals obtained using effluent recycle are much greater for low organic loadings than was obtained in other tests. The higher removals obtained at low organic loadings corresponded to the highest recycle ratios.

These results show that optimum conditions for biological activity in the trickling filter may be produced by recycling effluent. On the other hand, changes in operational conditions to optimize biological activity in the filters also result in a decrease in the effectiveness of the other principal BOD removal mechanism, i.e. air stripping. The results in Figure 18 show that increasing the recirculation ratio to approximately six greatly increased removals which can be achieved in the trickling filter. However, increasing the recycle ratios above six had very little additional effect on the improvement of removal efficiency.

Oxygen Uptake Measurements

The oxygen uptake rate is illustrated as a function of applied organic load in Figure 19. These data show that oxygen use in the raw waste tests increased slightly with applied organic load, but that the absolute values were significantly lower than in the dilution and recycle studies. Oxygen utilization in dilution tests was relatively constant over the range of loadings applied. In effluent recycle tests, oxygen utilization was highest at low

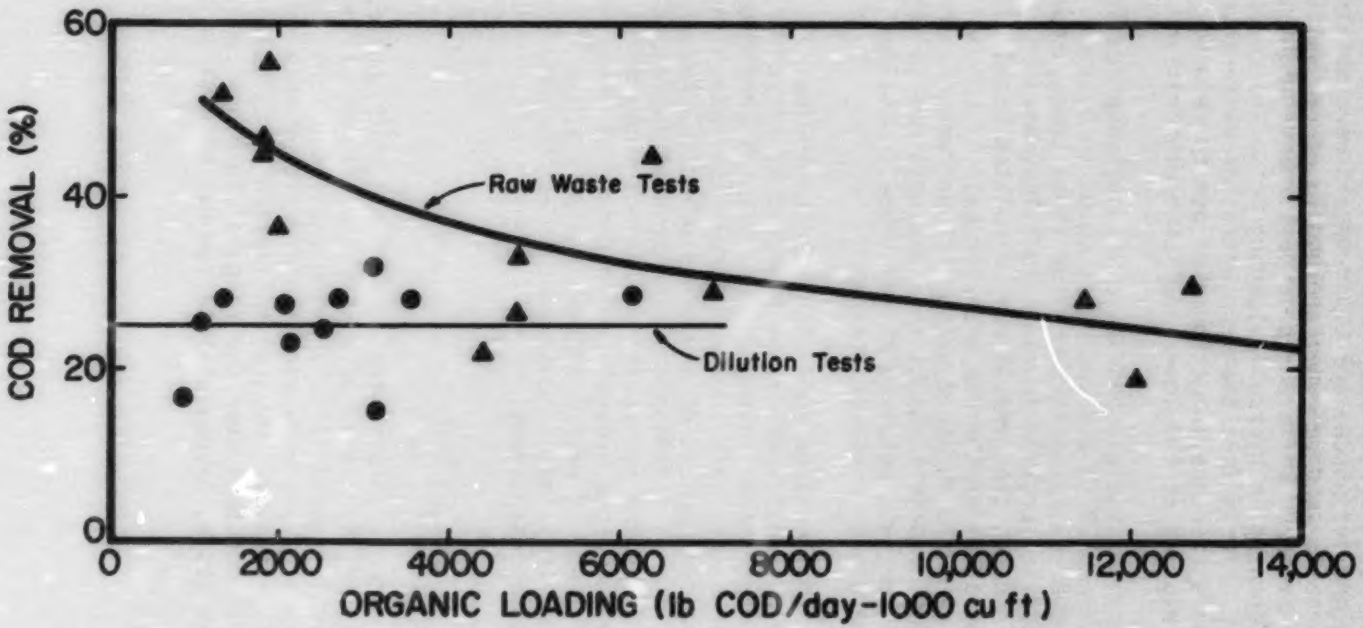


FIG. 16 .VARIATION OF REMOVAL EFFLUENT WITH APPLIED COD LOAD

(kg/cu m-day = lb/1000 cu ft-day x 0.016)

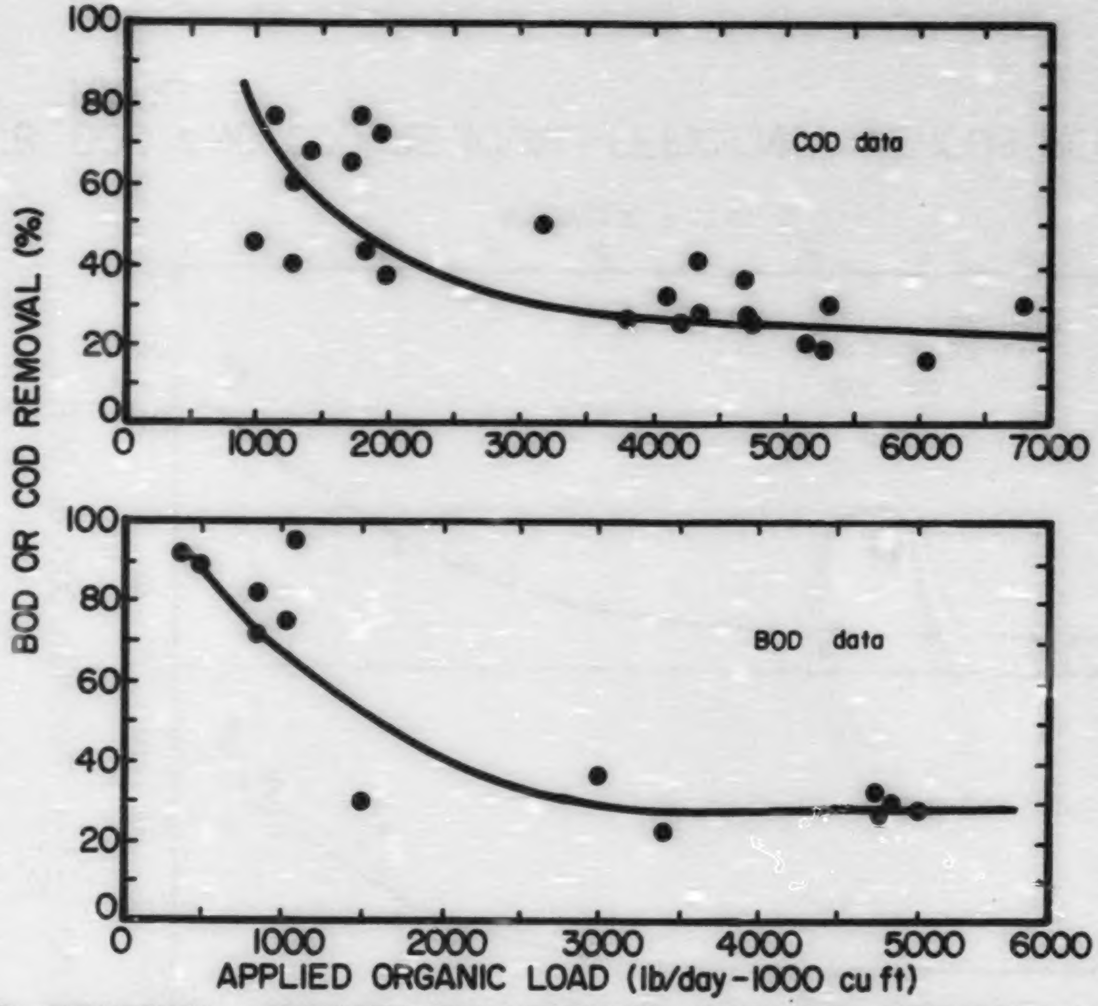


FIG. 17 .REMOVAL EFFICIENCY VERSUS APPLIED ORGANIC LOADING IN THE RECIRCULATION STUDY

(kg/cu m-day = lb/1000 cu ft-day x 0.016)

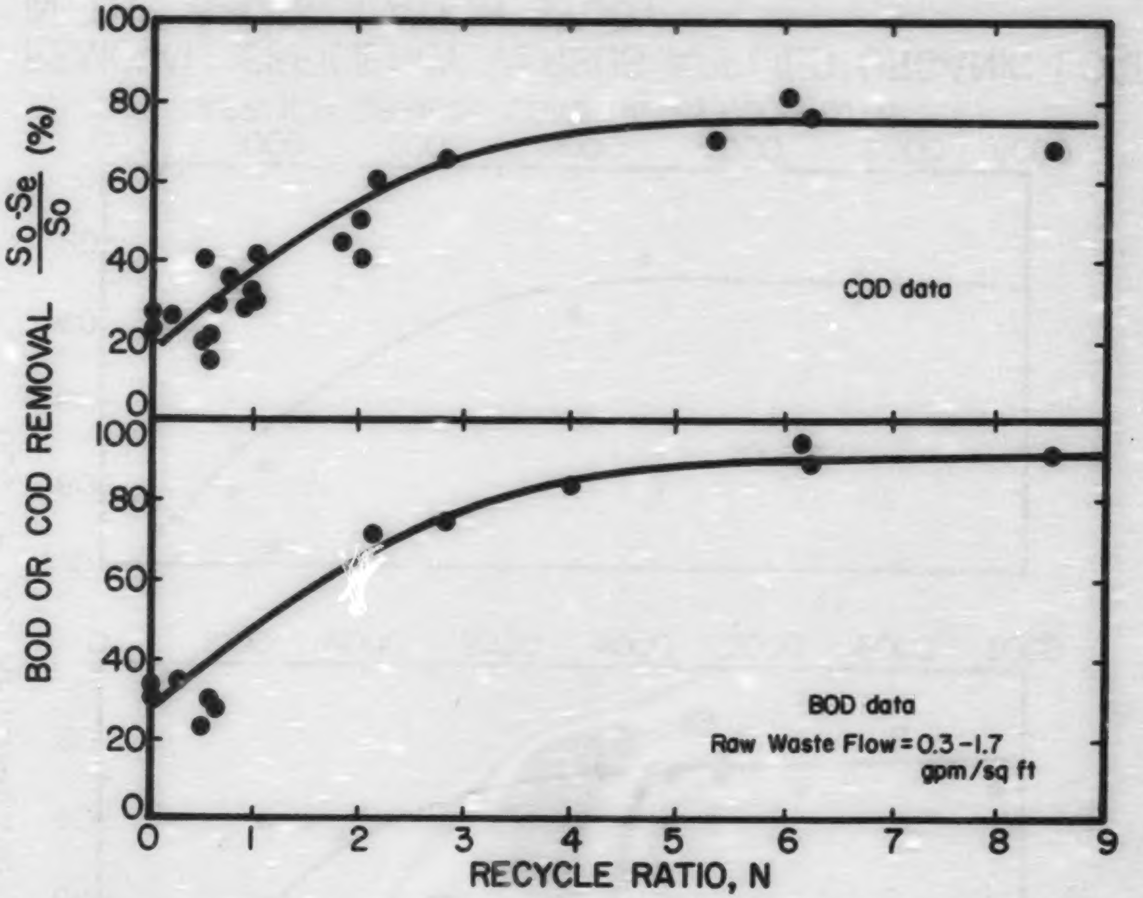


FIG. 18 . BOD AND COD REMOVAL EFFICIENCY VERSUS RECYCLE RATIO

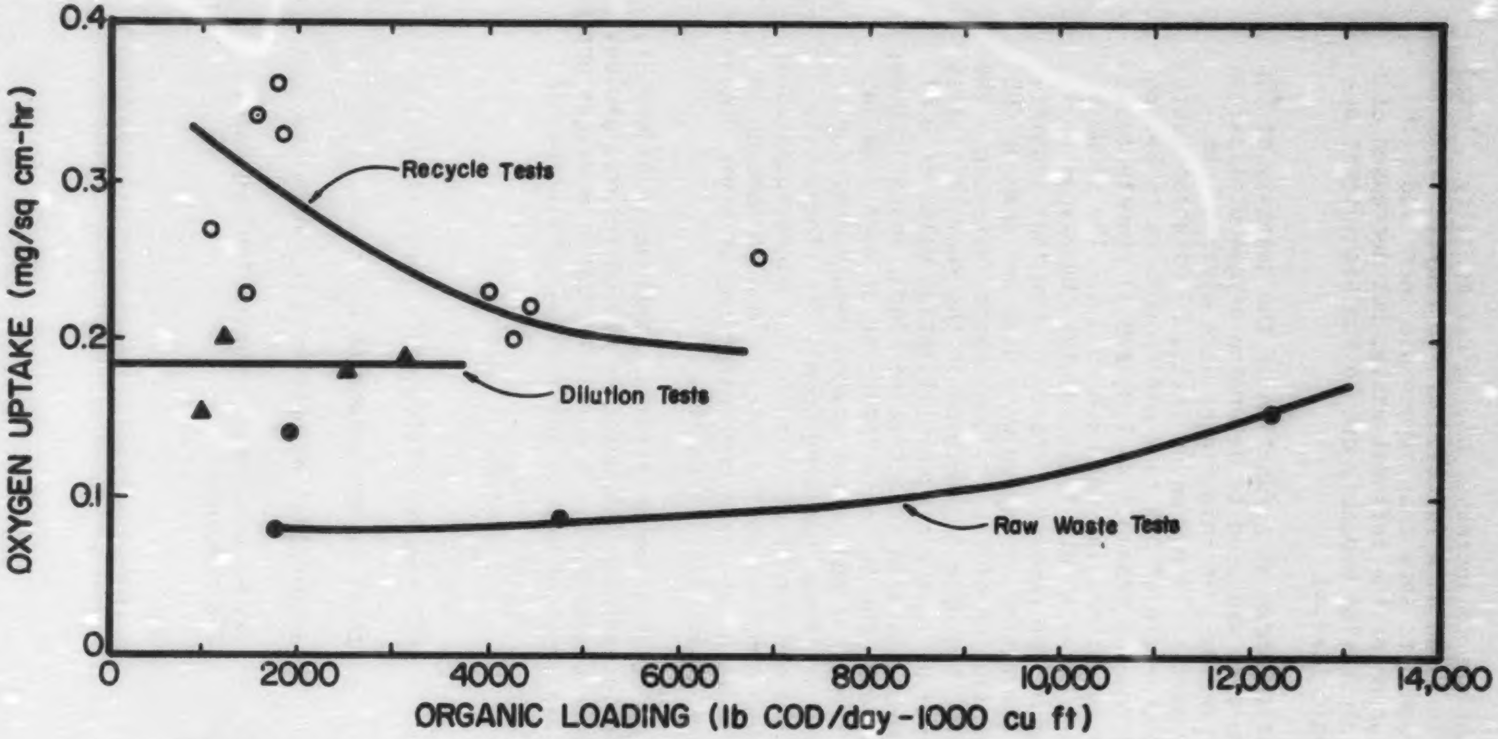


FIG. 19 .EFFECTS OF ORGANIC LOADING ON OXYGEN UPTAKE RATE
(kg/cu m-day = lb/1000 cu ft-day x 0.016)

organic loadings due to increased biological activity. Total oxygen uptake is shown as a function of the recycle ratio in Figure 20. These data show that oxygen use increased with increasing dilution of the influent waste, but decreased at very high recycle ratios probably due to the significant decrease in organic loading.

Oxygen utilization shown as a function of the quantity of COD removed in Figure 21 was used to determine oxygen utilization coefficients. The data in this figure were divided into three separate parts: 1) raw waste tests, 2) dilution tests having a dilution ratio of 0.5 to 2 and recycle tests having recycle ratios ranging from 0.2 to 1.9, and 3) recycle tests having recycle ratios ranging from 2.8 to 8.4. The slope of the line defined by each set of data yields the oxygen utilization per gram of COD removed. The data show that significantly higher quantities of oxygen were used per gram of COD removed as the dilution of the influent waste increased. The value for a' of 0.6 g O_2/g COD removed in the recycle studies was almost identical to the oxygen utilization value of 0.67 g O_2/g COD removed measured in activated sludge investigations using air aeration. The ordinate intercept of each line in this figure defines the endogenous oxygen requirement for microorganisms under each operating condition. These data indicate that the endogenous activity of the microorganisms is inhibited by high organic loading rates. This phenomenon is probably the result of the high concentrations or organic constituents at low waste dilution ratios.

Using the oxygen utilization coefficient obtained in Figure 21, it was possible to estimate the quantity of COD removed biologically from oxygen uptake data. Removals of COD from oxygen uptake data were calculated according to the following relationship:

$$(S_r)_B = (\Delta O_2 - b')/a'$$

where $(S_r)_B$ = COD removed biologically (g/sq cm-day)

ΔO_2 = total oxygen uptake (g O_2/sq cm-day)

b' = auto-oxidation uptake (g O_2/sq cm-day)

a' = oxygen utilization coefficient (g O_2/g COD removed)

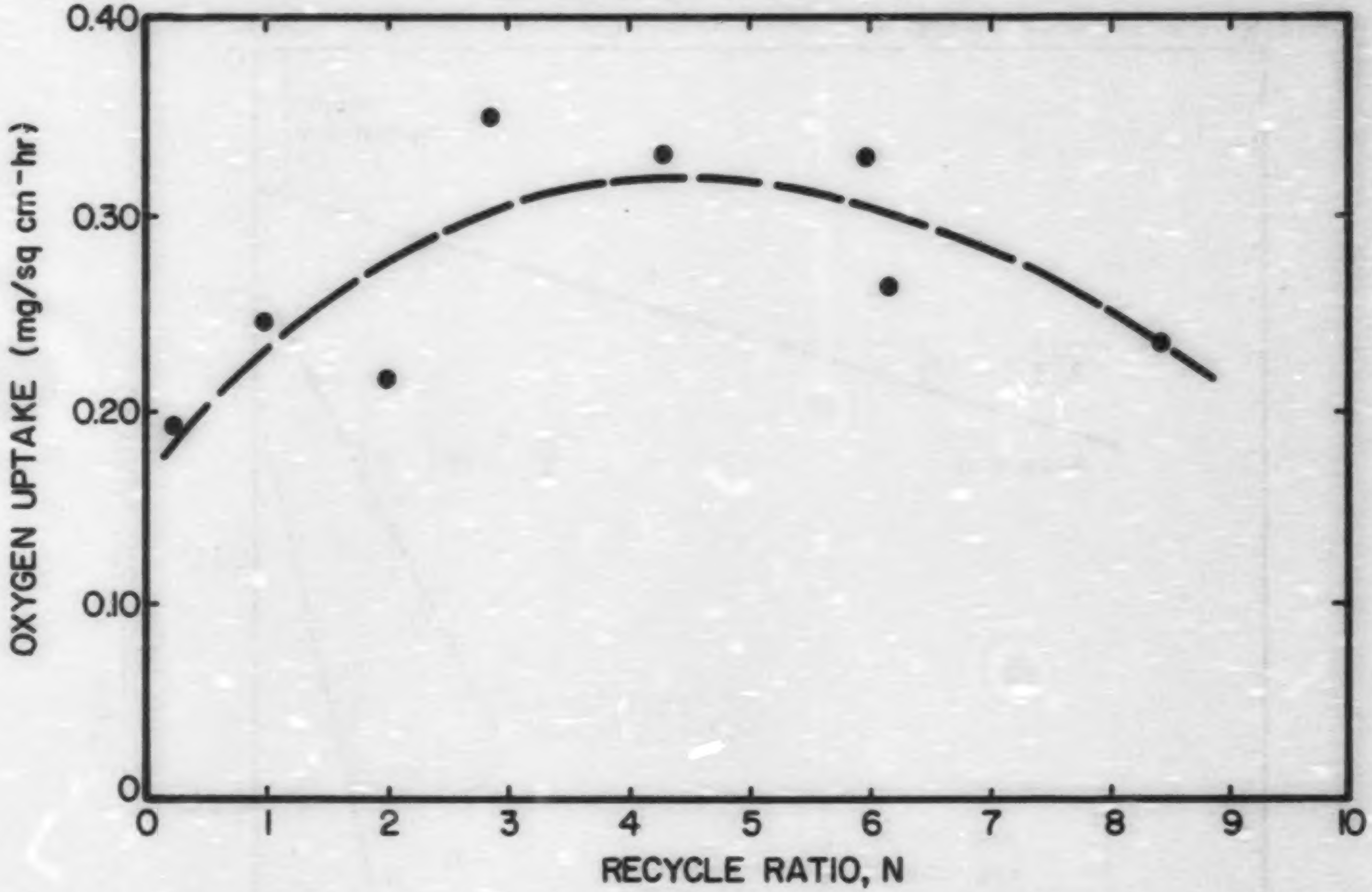


FIG.20.OXYGEN UPTAKE VERSUS RECIRCULATION RATIO

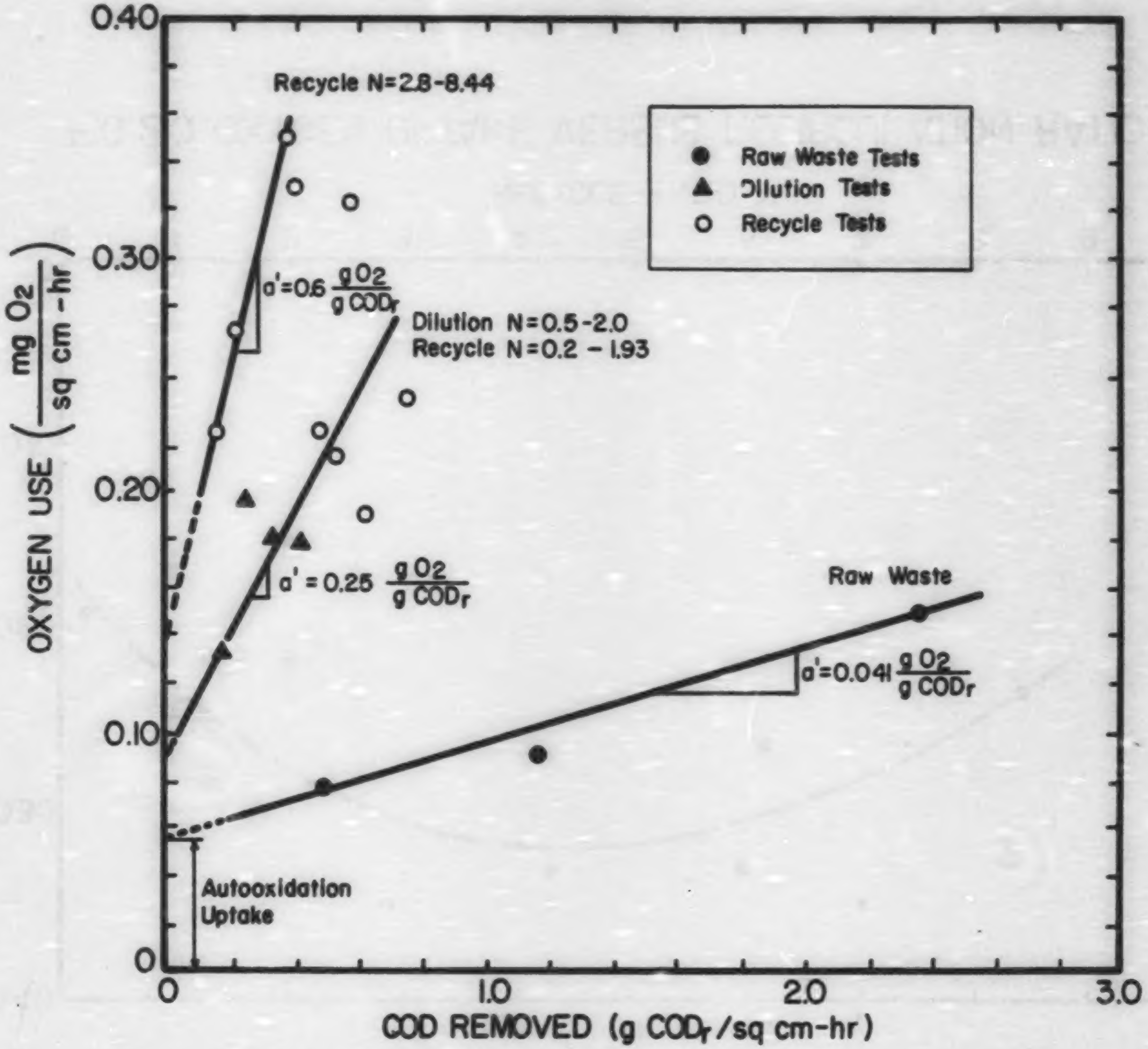


FIG. 21 .OXYGEN UTILIZATION IN THE MICROTOWER

The calculated COD removed according to oxygen uptake data, $(S_r)_B$, is compared to the measured COD removal, $(S_r)_T$, as a function of the applied COD concentration, S_a, in Figure 22. These data show that the fraction of COD removal accounted for by biological mechanisms increased significantly as the applied COD concentration to the filter decreased. At high applied COD concentrations these data indicate that only approximately 10 percent COD removal was due to biological mechanisms. For high influent COD concentrations, this value was significantly lower than that estimated for the full-scale treatment system. However, the results for the microtower were obtained from a single-stage trickling filter system, while oxygen uptake values discussed for full-scale system were total values for all six stages.

PILOT TRICKLING FILTER INVESTIGATIONS

A two-stage pilot trickling filter consisting of two 20-ft (6 m) towers having cross-sectional areas of 4 sq ft (1.4 sq m) were constructed to determine the effect of certain operational changes on the full-scale trickling filter system. The objective of these investigations was to determine the operating characteristics of the pilot system at various recirculation ratios and to determine the degree of BOD removal by air stripping occurring in the first part of the full-scale system. The pilot trickling filter was constructed so that equalized raw waste from the riser to the first tower of the full-scale system could be used as feed for the pilot system. Considerable difficulty was experienced in generating an adequate biological slime in these towers. Growth in the pilot system was very similar to the small amount of growth observed in the full-scale system. However, no difficulty was encountered in obtaining an adequate growth of microorganisms in the microtower. The difference might have been due either to the lesser degree of equalization provided in the pilot and full-scale systems or to differences in the characteristics of the plastic medium used. Pieces of wood placed at mid-depth in the second pilot tower showed a somewhat greater growth of slime than did the plastic medium, but the growth was still less luxuriant than would be expected for the organic loading rates which were used in this system. During the months of July and August, it appeared that the maximum amount of growth obtainable had been generated into pilot systems. Therefore, biological studies were initiated at that time. Prior to growing a biological slime in the pilot system, removals across the two pilot towers were monitored

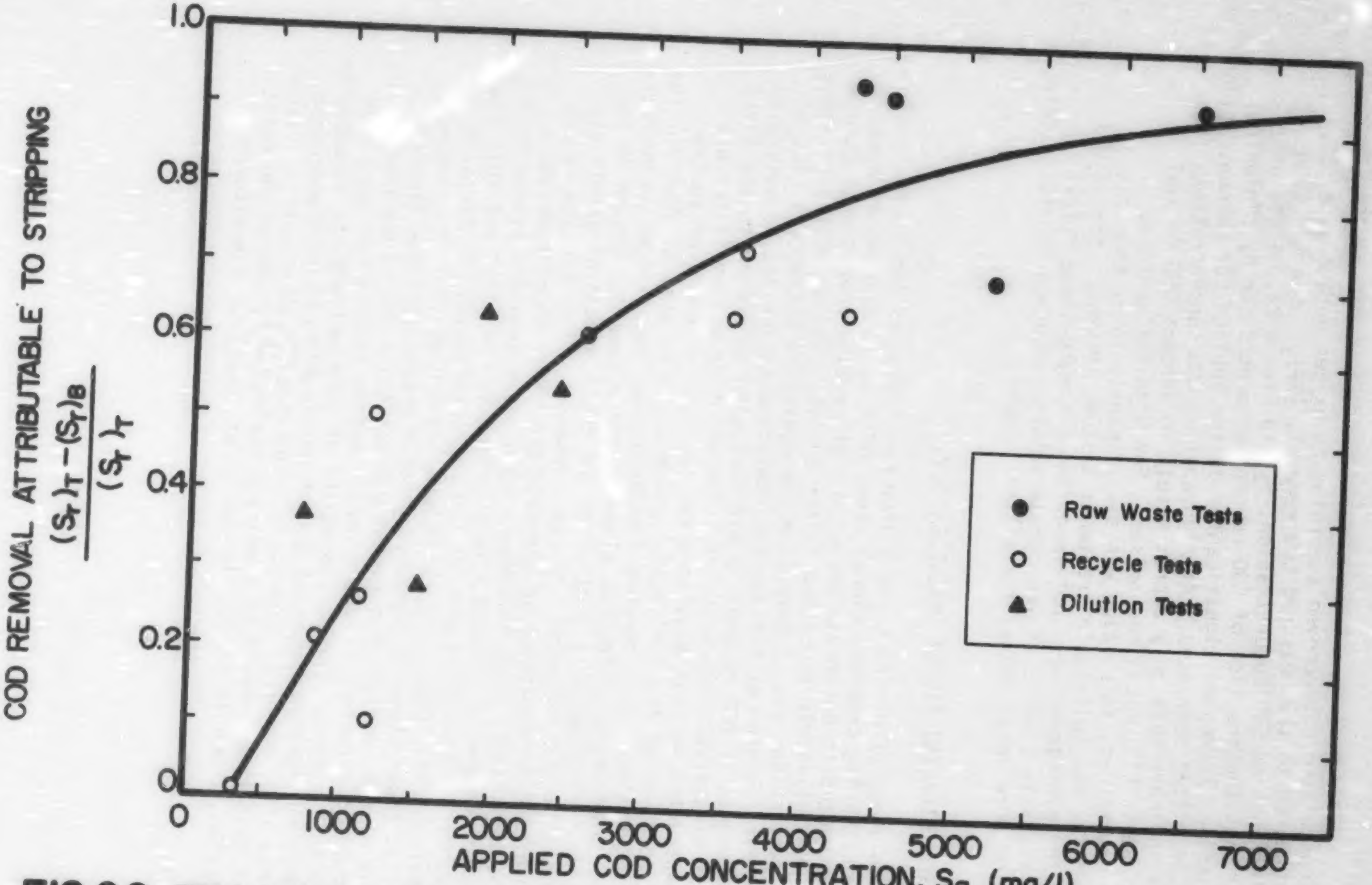


FIG. 22. FRACTION OF COD REMOVAL DUE TO OXYGEN UPTAKE IN THE MICROTOWER

in order to determine the amount of stripping occurring in the first part of the full-scale system. The results of these investigations are shown in Table 3. The average removal achieved over this period was significantly greater for Tower 1 than Tower 2 indicating that most volatile substances were removed through the first tower, while additional volatile materials were removed in the second tower, but to a lesser degree. Total carbon removals averaged 9.6 and 4.7 percent for Towers 1 and 2, respectively. Corresponding total carbon removals for days on which data were available for both the pilot and full-scale systems were 6.8 and 7.1 percent for the pilot system and 6.8 and 7.7 percent for the first two towers in the full-scale system. These data indicate that practically all removal of COD in the first two towers was due to the air stripping mechanisms.

Recycle investigations in the single-stage microtower indicated that approximately 85 to 90 percent BOD removal could be achieved at recycle ratios greater than 6.0. While this degree of recycle would not be possible in the full-scale system, further tests in the pilot towers were undertaken to determine removals which could be obtained with recycle ratios ranging between 1.0 and 3.0. The effect of recycle on COD removal in the pilot filter system is shown in Figure 23. The recycle data indicate that the performance of the full-scale trickling filter could be improved by adding recycle capabilities. Organic loading rates to the first tower for data shown in this figure were 915, 984, 770, and 152 lb COD/day-1,000 cu ft (1.47, 1.58, 1.23 and 0.24 kg COD/day-cu m) for recycle ratios of 1.0, 1.4, 1.67, and 3.0, respectively.

Because the strength of waste was somewhat lower than average during the period of these studies, organic loads applied to the filters were less than would be anticipated for the full-scale treatment facility. The variation of COD removal with organic loading is presented in Figure 24. These results show that the major fraction of COD removal occurred through the first tower indicating that the rate of degradation will decrease through the system. This might be due to stripping effects or to the removal of more readily degradable substances in the first tower.

TABLE 3

STRIPPING RESULTS FROM PILOT TRICKLING FILTER

Date (1973)	COD (mg/l)			TSC (mg/l)		
	Influent	Tower 1 Eff.	Tower 2 Eff.	Influent	Tower 1 Eff.	Tower 2 Fff.
3/12	-	-	-	1450	1300	1230
3/14	6720	5720	5080	1650	1690	1500
3/15	-	-	-	1640	1420	1420
3/16	5920	5670	4800	1530	1600	1295
3/20	-	-	-	875	450	470
3/21	4200	3400	3800	1115	1025	1090
3/23	3460	3840	5080	-	-	-
3/24	-	-	-	1325	1160	1185
3/29	5040	4640	4520	1140	1140	1115
3/31	-	-	-	1160	1090	1115
4/2	2940	2240	2280	750	650	600
4/3	-	-	-	975	950	875
Average		11.9%	6.4%		9.6%	4.7%

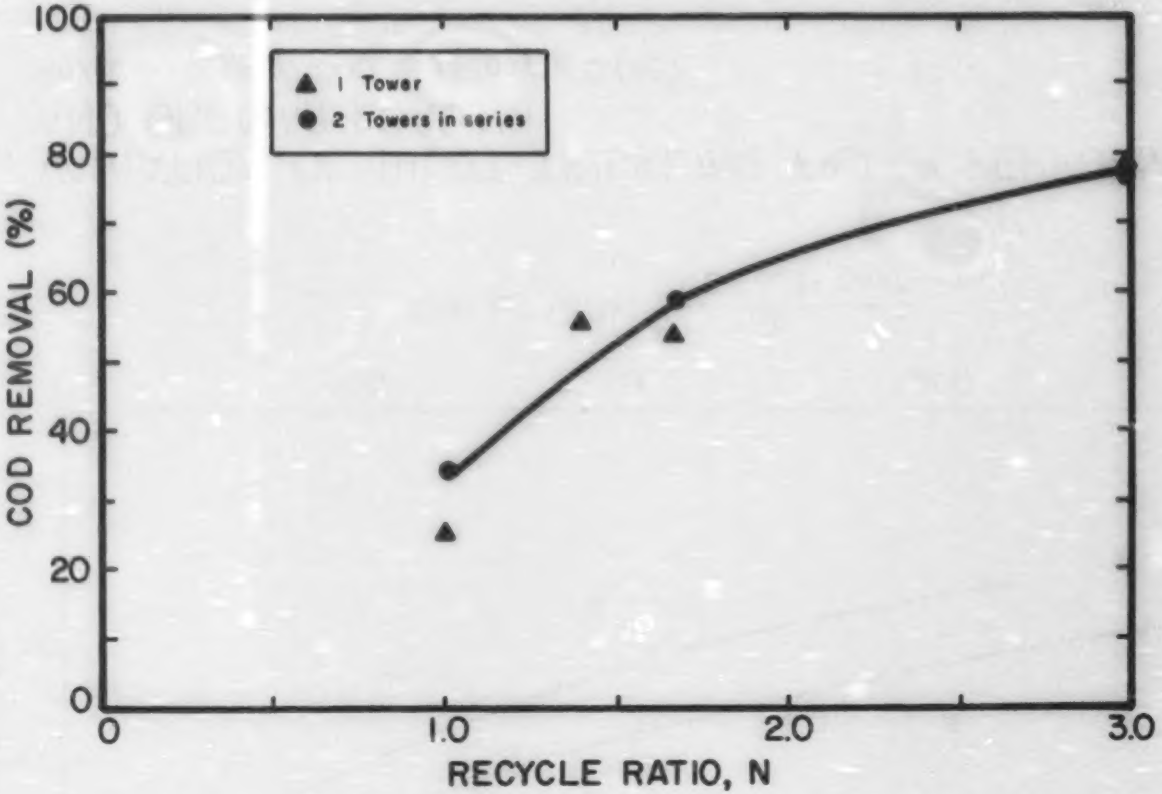


FIG. 23. EFFECT OF RECYCLE ON PILOT TRICKLING FILTER PERFORMANCE

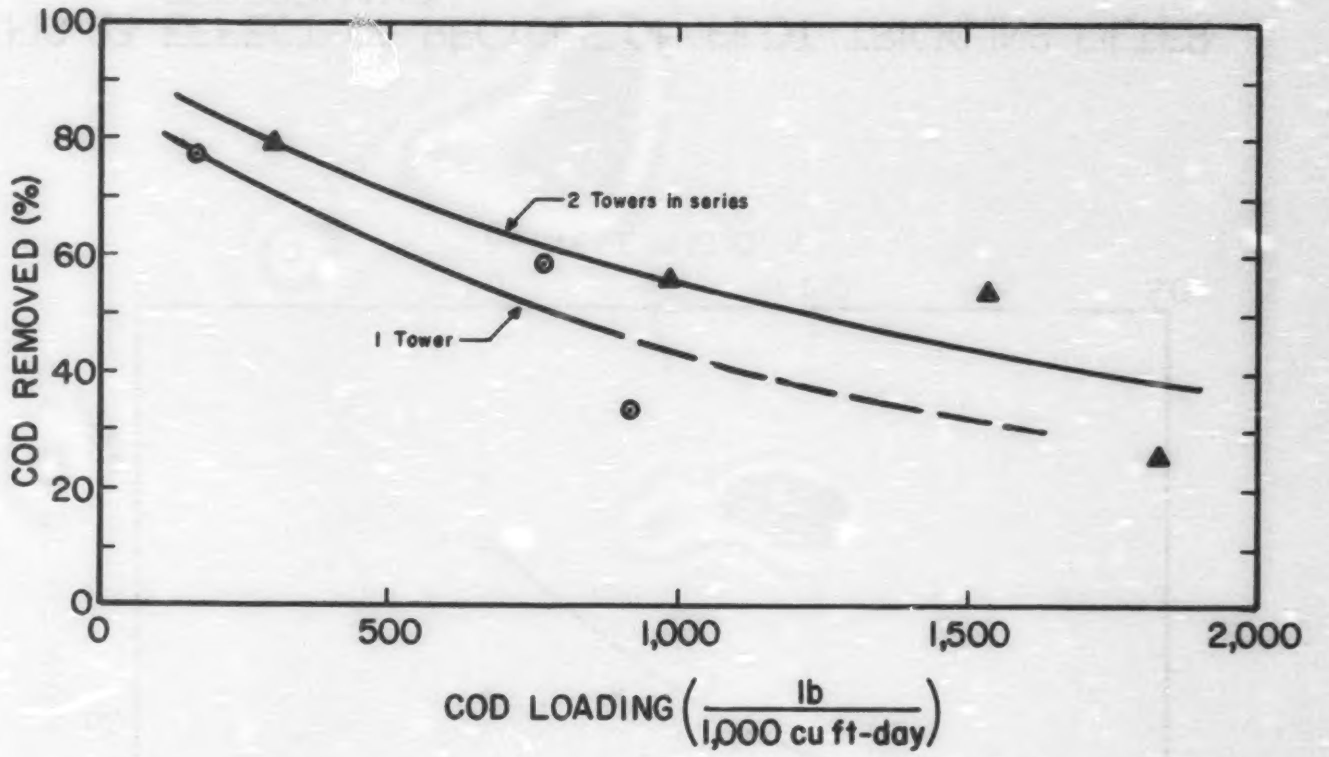


FIG. 24. VARIATION OF PILOT TRICKLING FILTER PERFORMANCE AND ORGANIC LOADING

(kg/cu m-day = lb/1000 cu ft-day x 0.016)

SECTION VII

FULL-SCALE TRICKLING FILTER INVESTIGATIONS

Performance of the present trickling filter facility was monitored for a nine-month period beginning in November, 1972. Following modification of the system to provide effluent recycle, additional tests were made during April and May, 1974. In addition to daily measures of plant operations and treatment efficiency, data were collected to determine organic carbon removals and oxygen availability in each of the six trickling filter stages. Additional data taken to provide further insight regarding system behavior included oxygen uptake rates using biota from the trickling filter and analysis for individual substances present in the wastestream. From November, 1972 through May, 1973 all raw equalized waste and trickling filter effluent samples were composited continuously over 24-hr periods. During June and July, 1973 increased biological activity in the waste due to higher temperatures made it advisable to implement another sampling procedure. Over this time, four grab samples were taken during each 24-hr period and composited for one daily sample. While some samples of waste from individual towers were taken continuously over a 24-hr period, most analyses were made from grab samples taken on the morning of the day on which analyses were performed.

Due to several upsets in the neutralization system during the months of December and January, the trickling filters were reseeded twice during February using mixed liquor from a nearby municipal activated sludge plant. This procedure was undertaken to insure as uniform a growth of biota as possible for subsequent studies. However, after reseeding the system, no definite beneficial effects were observed. In February, 1973, a batch equalization procedure was implemented to assure greater control over the influent waste pH. Beginning in March, 1973, only one pH upset occurred during the subsequent five-month period.

This chapter contains all data concerning the operation of the full-scale treatment facility along with an interpretation of the findings as they pertain to the upgrading of the system to meet anticipated regulatory requirements. A summary of operational data is presented in Appendix B.

ORGANIC REMOVAL PERFORMANCE

Variation of Organic Loading and Removal Efficiency

Due to variation in waste organic strength, applied flow rate, and seasonal temperatures, several fluctuations in trickling filter performance were observed during the course of the investigation. The chronological variation of BOD removal in the treatment system over this period is presented in Figure 25. Performance of the system averaged for monthly periods beginning in November, 1972 are shown in Table 4. While it appeared during March and April that a significant improvement in trickling filter performance would be realized in warmer weather, the results for May and June show that another factor significantly reduced removals for these two months. Data presented in Figure 25 indicate that the increased BOD reductions (judged on a percentage basis) observed during March and April were due to the decreased influent waste load, since the quantity of BOD removed remained at approximately 100 lb/1,000 cu ft-day for January through June. Following that time, 200 gpm (0.75 cu m/min) of cooling water was removed from the plant process wastestream resulting in increased BOD concentrations beginning the last half of June. Organic loadings applied to the filter system (lb BOD/day-1,000 cu ft or kg BOD/day-cu m) were higher than average during June and July due to intensive production schedules.

During February, March, and April, investigations were also undertaken to determine the performance of the system at reduced loading rates. Therefore, during March and April, changes were made in several variables which are expected to have a positive effect on treatment plant performance. These included: 1) a decrease in the influent BOD concentration, 2) a decrease in the applied hydraulic and organic loadings, and 3) an increase in temperature. While it is not possible to precisely identify the effect of these changes on system performance, it appeared that decreases in hydraulic loading did not produce significant improvements in system performance.

The relationship of applied BOD load to the quantity removed at various hydraulic loading rates is illustrated in Figure 26. A line which best fits these data indicates that an average of 33 percent BOD was removed in the system. These data also indicate that increasing quantities of BOD were removed at organic loading

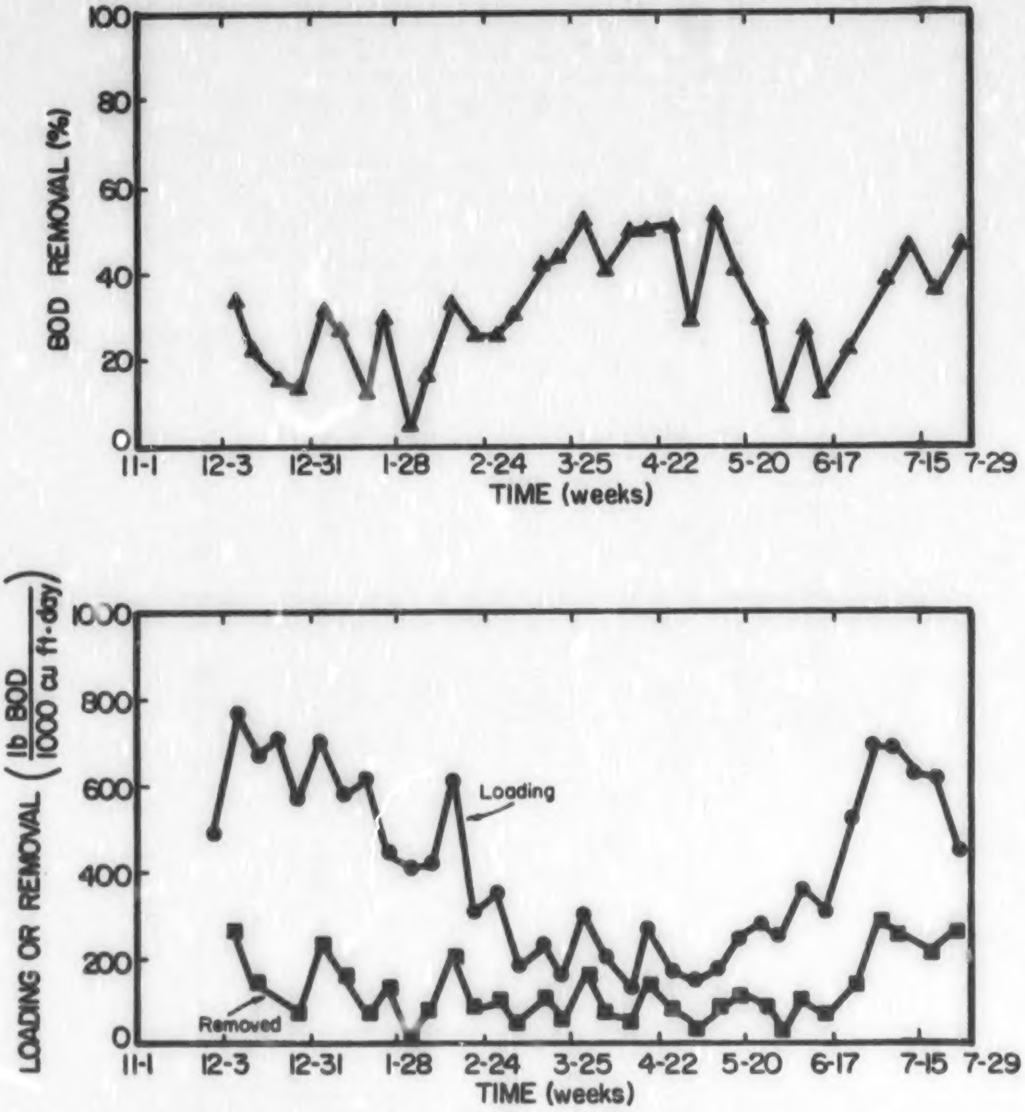


FIG. 25 .CHRONOLOGICAL VARIATION OF BOD REMOVAL
(kg/cu m-day = lb/1000 cu ft-day x 0.016)

TABLE 4

MONTHLY TRICKLING FILTER PERFORMANCE

Date (1972-73)	BOD (mg/l)				COD (mg/l)				TSC (mg/l)				Temperature (°F)			Avg Flow (mgd)[b]	System Loading[c] (lb organic/ 1000 cu ft-day)	
	Inf	Eff		Rem[a]	Inf	Eff		Rem[a]	Inf	Eff		Rem						
	Total	Total	Sol	(%)	Total	Total	Sol	(%)	Total	Total	Sol	(%)	Inf	Eff	Air		BOD	COD
Nov	3300	-	-	-	4279	2844	2841	33.5	1115	-	853	23.5	63	59	41	1.13	518	627
Dec	4600	3538	3310	23.1	5240	4278	3549	18.4	1586	-	1244	21.6	62	58	34	1.05	671	765
Jan	3204	2557	2098	20.2	5316	4142	3764	22.1	1402	-	1143	18.5	58	55	31	1.19	530	879
Feb	3574	2642	2548	26.1	5012	3637	3279	27.4	1407	-	1186	15.7	59	54	31	0.87	432	606
Mar	3002	1731	1566	42.3	4449	2741	2301	38.4	1257	-	810	35.6	62	60	43	0.56	223	346
Apr	1577	805	619	49.0	3002	1695	1368	43.5	828	-	493	40.5	65	63	49	0.89	195	371
May	1735	1147	1054	33.9	2869	2219	2026	22.6	800	-	663	17.1	70	64	57	0.99	239	395
Jun	3059	2365	2728	22.7	5029	3870	3576	23.0	1225	-	948	22.6	88	71	68	0.95	404	664
Jul	4345	2537	2287	41.6	6333	3881	3578	38.7	1376	-	1011	26.5	73	81	78	1.0	604	880

[a] Based on total influent and effluent values.

[b] Flow in cu m/day = mgd x 3.7.

[c] Loading to first filter is 6 times these values: loading in kg/cu m-day = lb/1000 cu ft-day x 0.016.

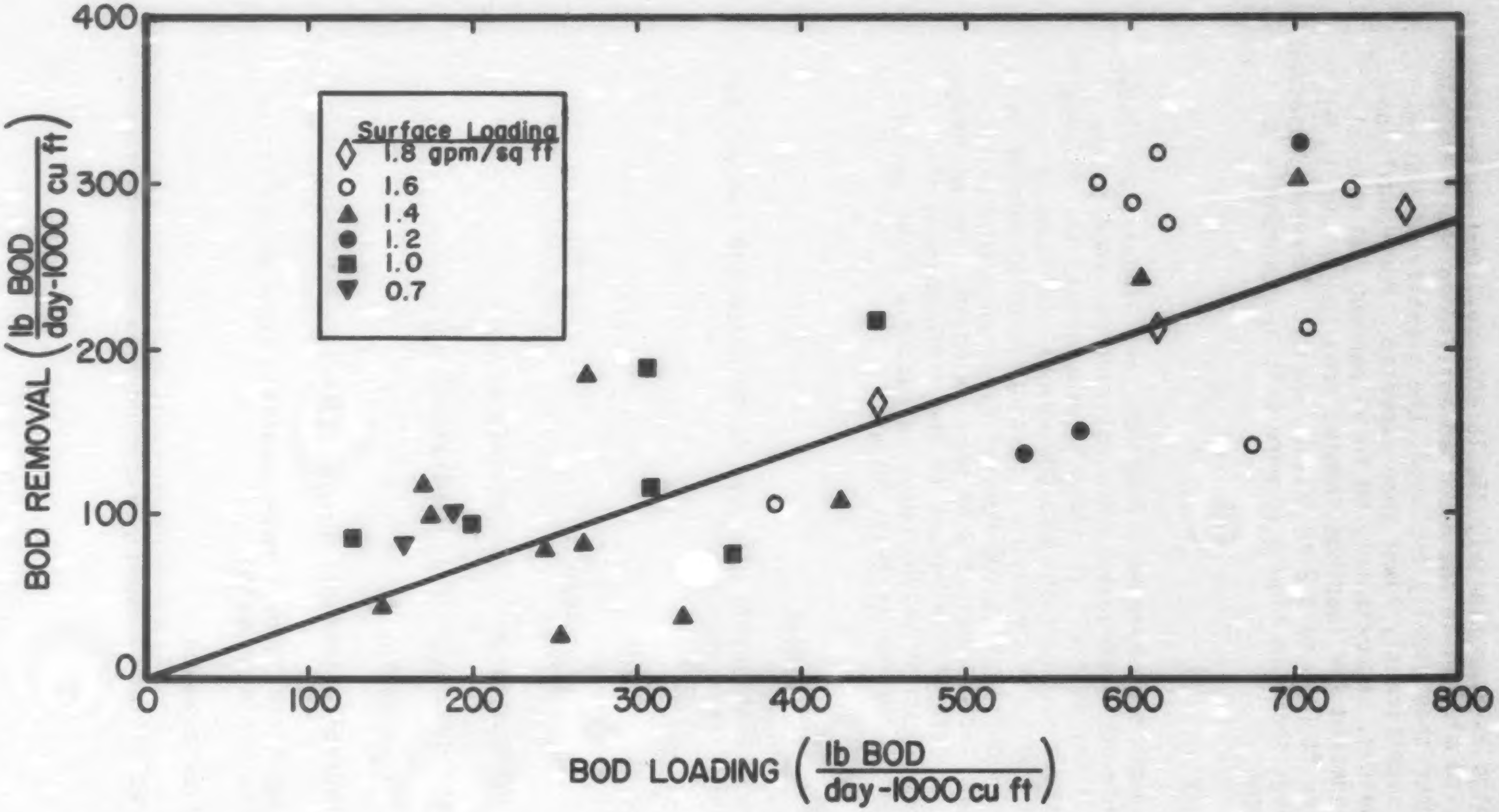


FIG. 26. RELATIONSHIP OF TRICKLING FILTER BOD LOADING AND REMOVAL

(kg/cu m-day = lb/1000 cu ft-day x 0.016; cu m/min-cu m = gpm/sq ft x 0.041) rates as high as approximately 750 lb BOD/day-1,000 cu ft (12 kg BOD/day-cu m). While the BOD load applied to these filters ranged higher than usually employed, the quantity of BOD removed was significantly lower than expected. For readily degradable wastes, approximately 80 to 90 percent BOD removal would be expected for loadings ranging from 100 to 200 lb BOD/day-1,000 cu ft (1.6 to 3.2 kg BOD/day-cu m). Removals observed in this range shown in Figure 26 were only approximately 30 to 70 percent.

Effect of pH

During the months of November, December, and January, problems were encountered in adequately neutralizing the feed to the trickling filter system. In order to demonstrate the adverse effect of these pH fluctuations, COD removal is shown as a function of the pH range during a particular 24-hr period in Figure 27. Reductions in COD removal efficiency were most noted for pH fluctuations of 3 or more pH units. The pH range identified in this figure refers to the difference in the minimum and maximum pH values observed during a 24-hr period which was sustained for at least 30 min.

BOD Removal Relationships

A first-order relationship was used to describe BOD removal in the trickling filter system:

$$\frac{S_e}{S_o} = \frac{e^{-KD/Q_A^n}}{(1 + N) - Ne^{-KD/Q_A^n}}$$

S_o = influent organic concentration, mg/l

S_e = effluent organic concentration, mg/l

D = filter depth, ft or m

Q_A = hydraulic loading, gpm/sq ft or cu m/min-sq m

K = substrate removal rate constant, $(\text{gpm/sq ft})^n$/ft or $(\text{cu m/min-sq m})^n$/m

n = flow exponent

N = recirculation ratio, recycled flow/raw waste flow

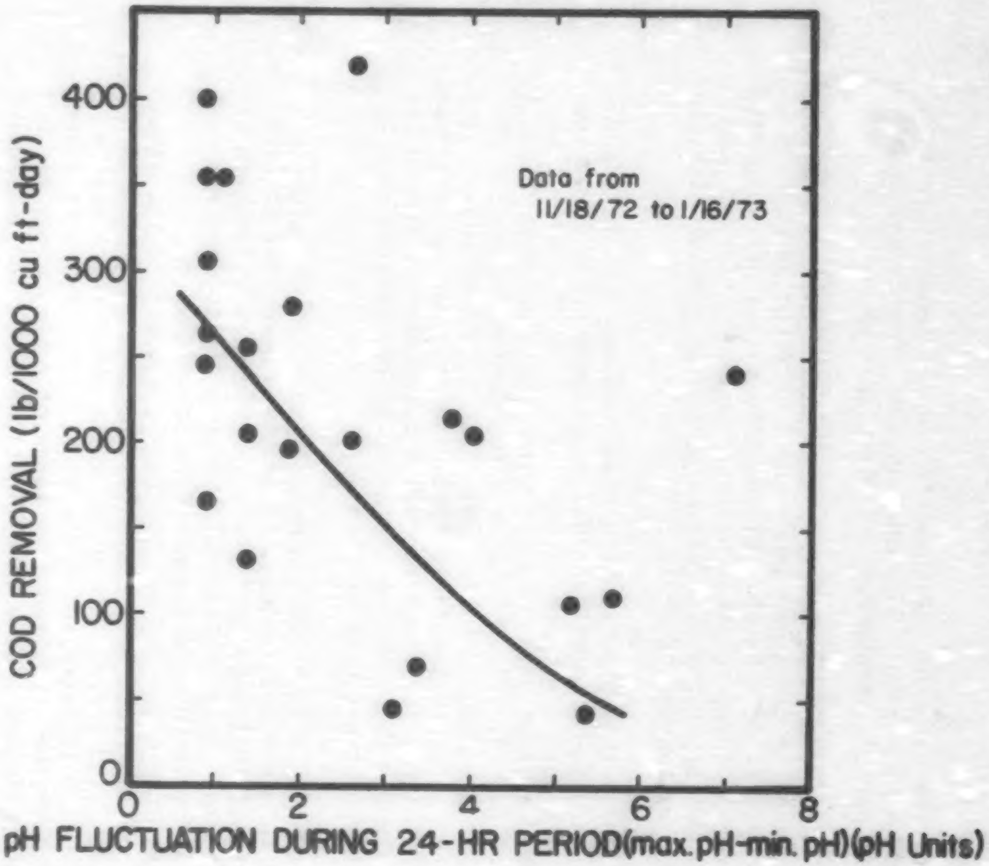


FIG. 27. EFFECT OF pH FLUCTUATION ON TRICKLING FILTER PERFORMANCE

(kg/cu m-day = lb/1000 cu ft-day x 0.016)

Preliminary data collected during this investigation indicated that the exponent, n, had a value of 1.0. Variation of the substrate removal rate constant, K, was described using the Streeter-Phelps relationship:

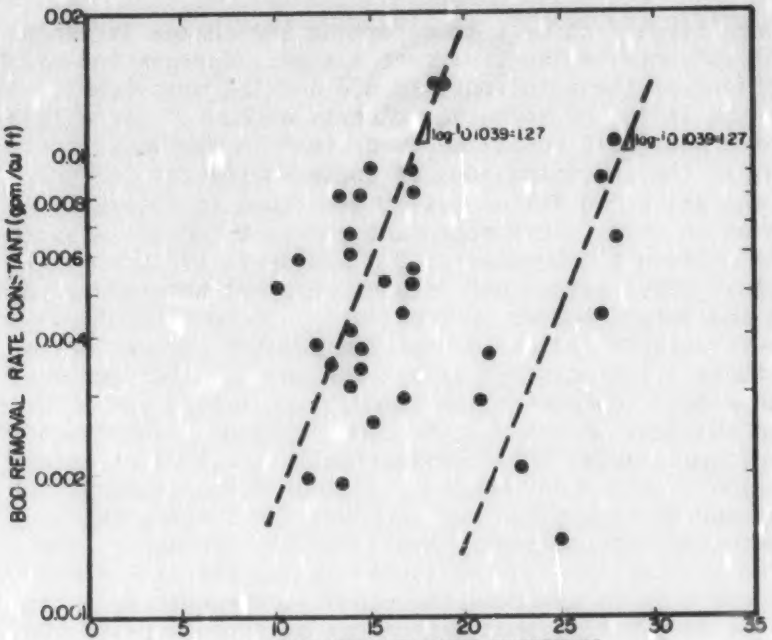
$$K_{T_1} = K_{T_2} \theta^{T_1 - T_2}$$

Variation of the value of K with waste and air temperature is illustrated in Figure 28. Data used in this figure were compiled for one-week periods throughout the investigation.

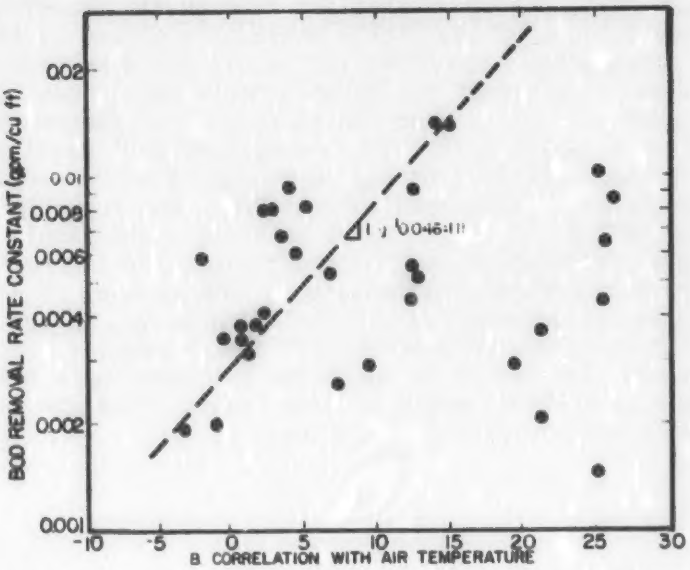
From these data, the temperature coefficient, θ, was determined to be 1.11 based on air temperature and 1.27 based on waste temperature. This is considerably higher than the values which have been reported in treating other wastes. However, there is some fundamental basis for the observation of higher θ values for complex organic wastewaters such as this one.

The variation of the data shown in Figure 28 was most likely due to changes in waste composition that might have occurred during the nine-month period of observation. The line of best fit in this figure was drawn including all points obtained from November, 1972 through May 26, 1973. Data obtained during the months of June and July are also shown in Figure 28, but there was a significant change in waste treatability during this period which cannot be attributed solely to temperature differences. Points for June and July are those having air temperatures greater than 19°C and waste temperatures greater than 20°C.

While temperature effects on biological processes are usually correlated to waste temperature, it can be argued that the temperature effect in trickling filters is primarily a function of air temperature. This is especially true of filter media which is not always covered by a laminar film of wastewater.
In this case, the organisms are constantly exposed to the atmosphere and not submerged in waste. For the data collected in this investigation, the correlation using waste temperature gave a value of θ which is extraordinarily high, while the correlation with air temperature resulted in a value of θ very close to that obtained in activated sludge investigations discussed subsequently in this report. At this time, no definitive conclusion may be reached because of the possible influence of other variables.



A. CORRELATION WITH WASTE TEMPERATURE



B CORRELATION WITH AIR TEMPERATURE

FIG. 28. TEMPERATURE DEPENDENCE OF THE BOD REMOVAL RATE CONSTANT

(cu m/min-cu m=gpm/cu ft × 0.13)

Fate of Solvents in the Trickling Filter System

Samples were taken each week to determine the change in concentration of solvents in the treatment system. Correlation of the concentrations of these solvents to BOD and TSC measurements made for the waste indicated that the solvents accounted for a large portion of the organic constituents present in the wastewater. Variations in the concentrations of these substances through the depth of the trickling filter system are shown in Figure 29. Values shown in this figure represent averages for all measurements made between mid-December, 1972 and July, 1973. In all cases, methyl ethyl ketone and toluene were not observed after the first and second towers, respectively. Detection of odors at the tower outlets indicated that the primary removal mechanism for these substances was by air stripping. The decrease in the isopropanol concentration through the system was accompanied by an increase in the acetone concentration. This suggested that the disappearance of IPA was partially a result of chemical oxidation to acetone. However, acetone concentrations increased by approximately 400 mg/l through the system, while IPA concentrations decreased by approximately 700 mg/l. The increase of acetone through the system is not completely understood; however, it is believed that this is due either to the formation of by-products from biological oxidation, or to the auto-oxidation of isopropanol through the system.

The stability of methanol concentrations through the towers is of significance because of the highly degradable nature of this substance. In a healthy biological system, it would be expected that methanol would be one of the first organic substances to be degraded. However, decreasing concentrations of isopropanol through the system indicate that the decrease of this readily degradable substance (IPA) might have been due to biodegradation. Because of the heavy organic loading and consequent poor BOD removal in the system the stability of methanol might simply be a result of the preference of the microorganisms for isopropanol. Generally, the data strongly indicate that some removal of organic constituents was accomplished by air stripping (e.g., toluene) and that, while there was evidence of biological removal of organic compounds, the extent to which the compounds were being removed was not as great as would be expected in a healthy biological system under normal loading conditions.

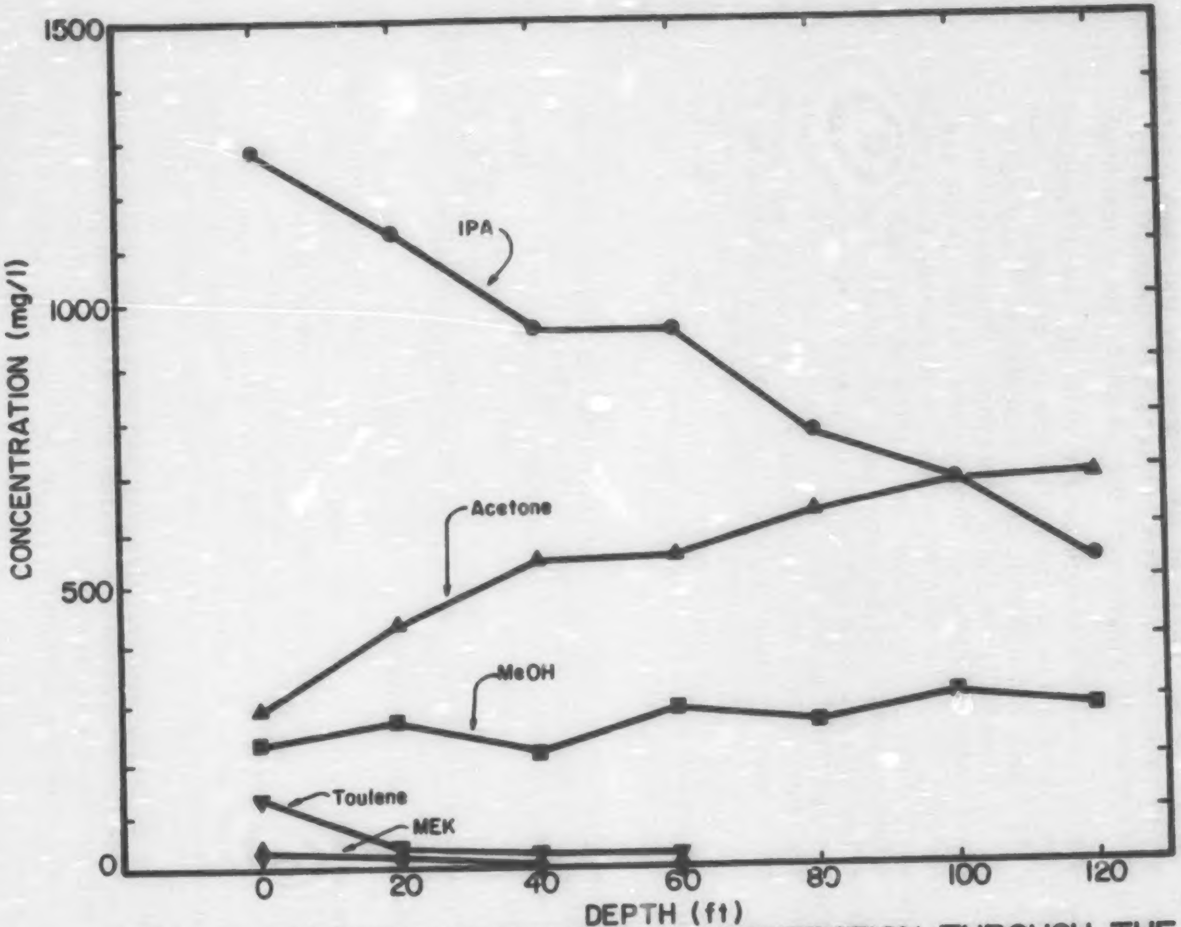


FIG. 29. VARIATION OF SOLVENT CONCENTRATION THROUGH THE TRICKLING FILTER SYSTEM

(m = ft x 3.3)

OXYGEN UTILIZATION

Oxygen uptake measurements in the full-scale trickling filter system were made to determine the effect of various operating variables on the oxygen uptake rate of the biota and as a means of elucidating the mechanism of BOD removal in the tower system. Oxygen uptake rates for trickling filters are conventionally expressed as grams of oxygen used per sq cm of media per hour.

In this application, the oxygen uptake rate will be a function of the quantity of biota on the media sample, the organic load to which the biota are exposed, and the presence of any toxic or inhibitory substances in the wastestream. The variation of oxygen uptake rates as measured through the filter system are shown in Figure 30. The date on which the measurement was made and the waste flow to the system are indicated for each measurement. For most tests, uptake rates were relatively low through the first two filters followed by significantly increasing values through the rest of the system. These results indicate inhibition of the biota in the first two towers followed by decreasing inhibition at greater tower depths. This inhibition might have been due either to pH and organic loading shocks felt most significantly in the first two towers or to the presence of inhibitory or organic compounds which are removed through the system either by air stripping or by biological oxidation.

While these results indicate significantly increased oxygen uptake levels at reduced waste application rates, it should be noted that these tests were made during the months of January, February, and March when flows were 1,000, 500, and 350 gpm (3.8, 1.9, 1.3 cu m/min), respectively. Thus, differences in temperature may have had some influence on these results.

Additional oxygen uptake values determined during the months of April and June are also shown in Figure 30. The uptake relationship measured on April 26 showed one of the highest levels of activity measured during the course of this investigation and corresponded to the time period in which the organic applied to the towers was the lowest during the 9-mo study. Oxygen uptake rates measured June 14, when BOD removal performance of the filters was generally poor, showed significantly less activity.

Examination of biota samples used for oxygen uptake analyses showed a significant increase of slime thickness through the treatment system during the spring months. Slime thickness in Towers 1 and 2 were approximately 1 mm or less with thickness

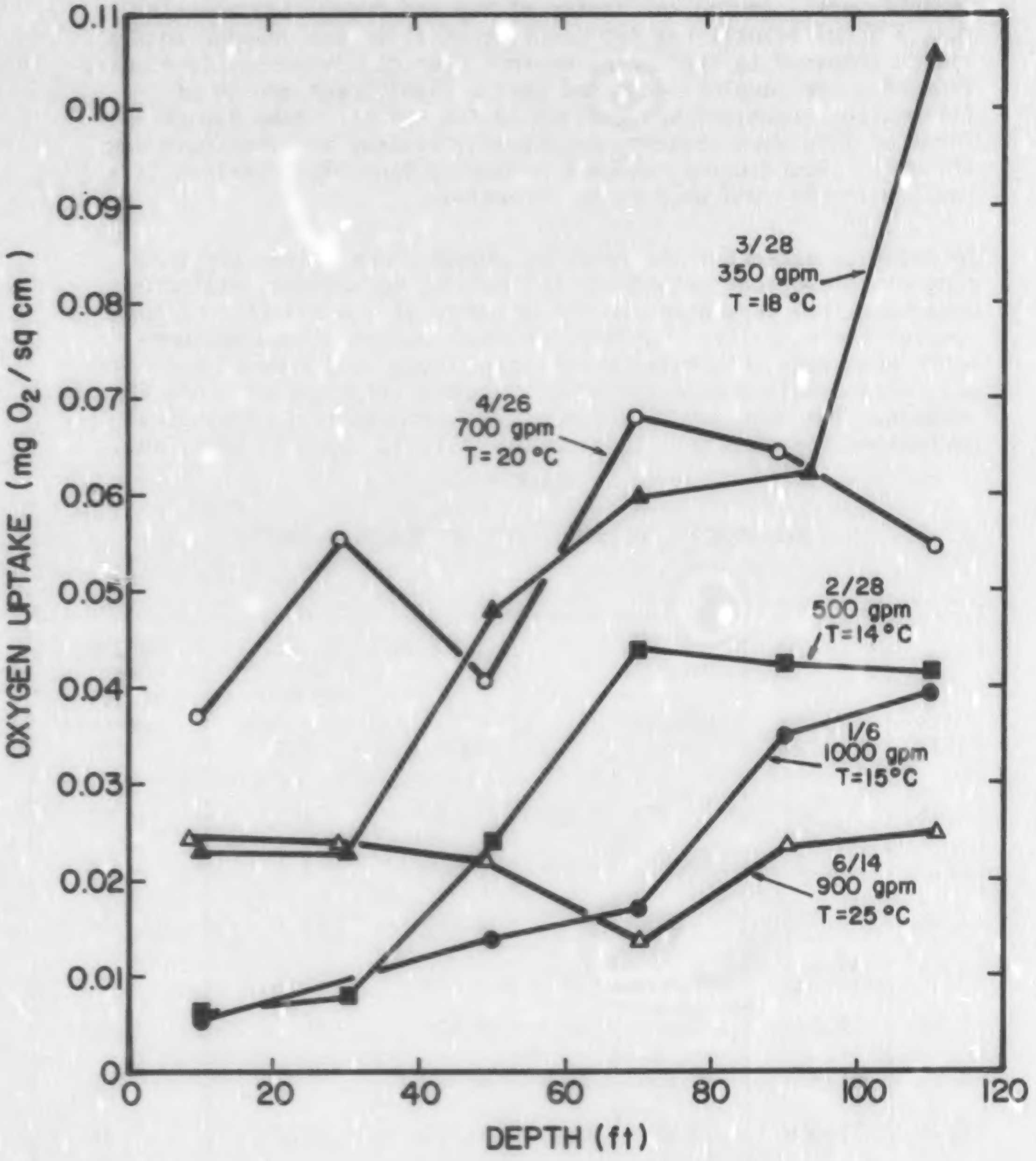


FIG.30. OXYGEN UPTAKE IN FULL-SCALE FILTERS
(cu m/min = gpm x 3.79 x 10^{-3}; m = ft x 3.3)

increasing in subsequent towers to approximately 2 mm in the last three towers. During the months of May and June, it appeared that a significantly smaller quantity of slime was present in the system compared to the spring months. Periodic microscopic examination of slime samples indicated that a significant growth of filamentous organisms was present in the system. Some higher forms of life were present, principally stalked and free-swimming ciliates. Round worms, which are usually found in trickling filters, were observed only on one occasion.

In order to determine the relative significance of the air stripping and biological methods as BOD removal mechanisms, estimates were made from oxygen uptake measurements of the quantity of BOD removed biologically. For days on which oxygen uptake measurements were made, estimates of the quantity of BOD removed biologically were obtained by assuming an oxygen utilization of 1 g/g BOD removed. The quantity of BOD removal attributable to biological mechanisms for each test is shown in Table 5. On most occasions,

TABLE 5

BOD REMOVAL ATTRIBUTABLE TO OXYGEN UPTAKE

Date (1973)	BOD applied (lb/day)[a]	BOD removed		BOD Removal Due to O_2 Uptake		
		(lb/day)[a]	(%)	(lb/day)[a]	Fraction of BOD applied (%)	Fraction of all BOD removed (%)
1/16	35,340	7,070	20	1,490	4.2	21
2/28	24,020	1,800	7.5	1,780	7.4	99
3/28	18,310	6,000	32.8	3,970	21.7	66
4/18	23,320	7,140	30.6	5,750	24.6	80
4/26	12,610	8,080	64.1	3,920	31.1	48
6/1	21,350	6,960	32.6	6,980	32.7	100
6/14	30,080	3,370	11.2	3,250	10.8	96
6/22	52,780	18,580	35.2	8,700	16.5	47

[a] (kg/day = lb/day x 0.45).

the oxygen uptake rate for the waste alone was determined at two or three depths in the tower system. For measurements made through the month of May, the uptake of the control waste sample was very small and did not significantly affect the value measured for the uptake of the organisms attached to the filter medium. For measurements made during June and July, it was necessary to determine the oxygen uptake rate of the waste separately and correct the value obtained for the waste plus the medium sample for the uptake of the waste alone. Visual observations indicated that a biological growth established itself in the sumps beneath each tower during the summer months. Because these organisms also contributed to the biological removal of organic constituents, an estimate of the removals accomplished by these dispersed organisms were included in estimates of biological BOD removal.

These results show that significantly greater BOD removals were attributable to biological mechanisms during the month of April. Smaller biological removals of BOD during the months of January and February were probably due to lower temperatures and inadequate waste neutralization. The relatively poor performance during the months of June and July cannot be explained. While the quantity of BOD removed biologically varied considerably during the 6-mo period of observation, it appeared that biological mechanisms account for approximately 50 percent of the total BOD removed in the trickling filter system.

EFFECT OF FORCED DRAFT VENTILATION ON BOD REMOVAL

As a further indication of the relative significance of air stripping and biological mechanisms of BOD removal, tests were made to determine the effect of forced draft ventilation on removal efficiencies. For a one-week period during the month of May, dissolved oxygen and total carbon values were measured for the effluent from each tower with and without the forced air ventilation system running. After the fans had run throughout the night, samples were taken with all fans running each morning. Following the collection of samples, the blowers were turned off for several hours to allow the system to equilibrate to the new conditions; subsequently, samples were taken with the blower system not running.

The results of these tests are summarized in Table 6. The results show that while dissolved oxygen levels were lower in each tower when the fans were off, this level was not too low for aerobic biological activity. Total carbon removals for this system for the two conditions exhibited significant differences. For the

TABLE 6

EFFECT OF TOWER BLOWERS ON DISSOLVED OXYGEN LEVEL AND TOTAL CARBON REMOVAL[a]

Sample	Dissolved Oxygen (mg/1)		Temperature (°C)		Total soluble carbon	
	Fans on	Fans off	Fans on	Fans off	Fans on[b]	Fans off[c]
Influent	3.1	3.2	19.8	21.0	908	985
Tower 1	6.7	5.9	19.8	21.2	846	949
Tower 2	7.0	5.7	19.7	21.3	788	913
Tower 3	7.0	5.2	19.5	21.6	741	882
Tower 4	6.9	4.8	19.2	21.7	687	861
Tower 5	7.0	4.6	19.2	22.3	649	832
Tower 6	7.3	5.0	19.2	22.3	611	810

[a] Data taken daily over period from May 14 to May 19, 1973.

[b] Removal averaged 32.7 percent.

[c] Removal averaged 17.8 percent.

one-week period, total carbon removals during periods with forced draft ventilation averaged 32.7 percent, while the removal without forced ventilation was only 17.8 percent. Because the dissolved oxygen measurements indicated that D.O. was not limiting the biological performance of the system, it was concluded that the differences in removal observed with and without the blowers running can be attributed to air stripping of volatile components from the wastestream. While some air stripping probably occurred when the blowers were not running, it is felt that these reductions were minimal due to the great reduction of air flow through the system. While strictly quantitative judgments cannot be made on the basis of these data, it appears that biological and air stripping mechanisms were responsible for approximately the same degree of BOD removal, i.e., 50 percent of the total removal.

FULL-SCALE EFFLUENT RECYCLE INVESTIGATIONS

Results of pilot-scale trickling filter investigations conducted earlier in the investigation showed that effluent recycle was capable of improving treatment performance. Based on these favorable results, the trickling filter system was modified to provide continuous effluent recycle. In making this modification to the system, the six trickling filters were divided into two parallel systems, each containing three filters in series. Flow recorders and controllers were added so that both the raw feed rate and the recycle flow could be adjusted. Recycle flow consisted of unclarified effluent from each of the parallel systems.

Bank A (originally Towers 1, 2, and 3) was fed 700 gpm (2.6 cu m/min) of raw waste and 300 gpm (1.1 cu m/min) of recycle waste, while Bank B (originally Towers 4, 5, and 6) were fed 300 gpm (1.1 cu m/min) of raw waste and 700 gpm (2.6 cu m/min) recycle flow. Following stabilization of this period on April 1, 1974, data were collected to demonstrate the performance of this modified system.

A summary of data obtained for a 7-wk period is shown in Table 7. Weekly data are presented in Table B2 of Appendix B. During the recycle investigations, Bank A, operated at a recycle ratio of 0.43, achieved a BOD reduction of 22 percent which was somewhat less than the 1973 average of 32 percent for six filters in series. Bank B, operated at a recycle ratio of 2.33, achieved approximately 45 percent BOD reduction. For operation at identical hydraulic loading rates, it would be expected that the three filters in Bank A would achieve less BOD removal than six filters in series. The lower organic concentration in the first filter would probably reduce air stripping of organics but might not be significant

TABLE 7

RECYCLE DATA SUMMARY[a]

Bank	Recycle	Air temp. (°C)	BOD results				COD results			
			Inf. (mg/l)	Eff. (mg/l)	Reduction (%)	Applied load (lb/1000 cu ft-day)[b]	Inf. (mg/l)	Eff. (mg/l)	Reduction (%)	Applied load (lb/1000 cu ft-day)
A	0.43	11.4	2,203	1,711	22.3	820	3,925	3,330	15.1	1,500
B	2.33	11.4	2,203	1,216	44.8	605	3,925	2,683	31.6	1,220

[a] Data summarized for operation during 7-wk period from April 1 to May 19, 1974.

[b] Loading in kg/cu m-day = lb/1,000 cu ft-day x 0.016.

enough to enhance BOD removal. On the other hand, reductions in Bank B, operated at a recycle ratio of 2.33, were significantly greater than the 1973 average. This was perhaps due to the dilution of the influent organic concentrations and dampening of fluctuations in the organic load to the filter.

On a BOD basis, the total load applied to Bank A (including the BOD content of the recycle stream) was approximately 820 lb BOD/1,000 cu ft-day (13 kg/cu m-day), while that applied to Bank B was approximately 600 lb BOD/1,000 cu ft-day (9.6 kg/cu m-day). This is somewhat higher than the total load applied to a system without recycle. However, the load applied to the first tower in Bank B (approximately 1,800 lb BOD/1,000 cu ft-day or 28.8 kg/cu m-day) was significantly less than the load of approximately 3,000 lb BOD/1,000 cu ft-day (48 kg/cu m-day) applied to the first tower when six filters were operated in series.

It should be mentioned that the biota present in Bank B at the beginning of these tests was probably better developed than that present in Bank A. Operation of the recycle system over a longer period of time might result in greater BOD reductions at a recycle ratio of approximately 0.4 than those experienced in this investigation.

SECTION VIII

AUTOMATIC MONITORING

One objective of this grant was to evaluate the potential for controlling the multi-stage trickling filter system using automatic monitoring equipment. The methods to be evaluated included control of the raw waste feed rate using automatic total carbon measurements and control of the blower speed in the forced air ventilation system using dissolved oxygen measurements. Tests conducted during this investigation had the objective of demonstrating the applicability of automatic monitoring equipment in the trickling filter system. The investigation did not include control of the feed rate or blower speeds; however, a control system could be designed once the stability of monitoring systems has been established. During the fall and winter of 1973-1974, dissolved oxygen and total carbon monitoring equipment was installed to measure the contents of each sump and the clarifier effluent. A more complete description of these systems is presented in Section IV.

DISSOLVED OXYGEN MONITORING

Comparative dissolved oxygen analyses were made of measurements from the on-line system and a laboratory dissolved oxygen meter. For each comparison, grab samples were taken from the line feeding each D. O. probe. Simultaneous readings were made from the on-line D. O. indicator. The portable laboratory dissolved oxygen meter was taken to the field to provide a comparison for values measured using the on-stream system. This portable meter was calibrated in the laboratory prior to each use. A least squares analysis was performed on comparative analysis at each sampling point over a one-month period. Results of these analyses and average D. O. concentrations observed for both the laboratory and on-stream systems are summarized in Table 8. During this period routine maintenance in the system consisted of cleaning the probe assemblies at least once each week and recalibration of each probe approximately two times per week in order to maintain good agreement between the on-stream and laboratory analyzers.

The data presented in Table 8 show that the average D. O. differences between the two measurements ranged from 0.0 to 0.4 mg/l for

TABLE 8

COMPARATIVE DISSOLVED OXYGEN MEASUREMENTS

Sample location	Average D. O. concentration[a] (mg/l) lab meter	on-stream system
Tower 1	9.0	8.9
Tower 2	8.5	8.2
Tower 3	8.9	8.9
Tower 4	8.4	8.3
Tower 5	8.4	8.2
Tower 6	8.9	8.5
Clarifier	7.0	6.0

[a] Averages are from 14 comparative measurements made during a one-month period.

each of the towers and 1.0 mg/l for the clarifier probe. Considerable difficulty was encountered in obtaining proper operations of the clarifier probe. Shortly before investigations with the automatic monitoring system were initiated, chlorination of the effluent was begun by adding approximately 50 mg/l chlorine to the clarifier influent. As long as the clarifier probe remained immersed in the waste, readings recorded in the automatic monitoring system for this probe continually drifted downward. Upon changing the membrane of this probe and placing it in contact with tap water, readings made over a one-week period indicated that stable operation could be maintained with tap water. When the probe was returned to the wastewater system, the same lack of stability was noted and readings again drifted downward.

The least squares analysis of data for each sampling point indicated that somewhat lower readings were obtained in the on-stream system compared to laboratory measurements. A regression analysis of all data collected had a slope of 1.00 and y-intercept of -0.4 mg/l. This indicated that measurements using the laboratory meter were approximately 0.4 mg/l greater than measurements in the continuous-flow system.

Additional measurements were taken to determine the operation of the system when calibrated less frequently. While collecting data summarized in Table 8, each probe was recalibrated approximately two times per week. During a subsequent period, probes were calibrated approximately once each two weeks. Data collected with less frequent calibration showed that readings with the monitoring systems tended to drift downward. Comparative analyses made during this period showed that on-stream measurements averaged approximately 1.1 mg/l lower than equivalent measurements made using the laboratory analyzer for each of the tower probes. Dissolved oxygen concentrations indicated with the clarifier probe averaged approximately 3 mg/l less than comparative analysis made with the laboratory meter.

Observations made during this period showed the need for calibration of the system approximately twice each week. Experience indicated that if the probes were not cleaned or calibrated for a period of one week, the dissolved oxygen concentrations were approximately 20 percent lower than measurements made using the laboratory meter. Operationally, the system was well designed to allow for routine maintenance. The only maintenance required included cleaning and recalibration of each probe assembly periodically. To disassemble and clean the probe assembly, change the membrane, air calibrate the instrument, and reassemble the probe unit required approximately two hours per probe. Thus, approximately two days were required to clean and recalibrate the seven probes included in this system.

TOTAL CARBON MONITORING SYSTEM

During the initial one-month period of operation, distinct sample peaks were obtained only five days. Of the other days during this period, the instrument was not operational due to plugging of the sample prefiltration system (1 day), a leaking sample injection valve (16 days), and a plugged injection tube (2 days). During this test period the analyzer was not operational due to mechanical malfunctions 75 percent of the time. During the remaining 25 percent of the test period, values were recorded; however, most of these values did not agree well with laboratory measurements.

Required maintenance included replacement of the sample injection valve, packing of the reaction chamber with a new catalyst, replacement of the oven and mounting plate, cleaning of the flow regulator, and complete calibration of the instrument. The original catalyst used in the reaction chamber became poisoned soon after the instrument was placed into operation. Although a second type of catalyst was evaluated, the same problem recurred. It appeared that problems with plugging of the sample injection tube, poisoning of the catalyst, and leakage in a block unit used to meter samples into the reaction chamber were all caused by tar-like and other substances in this particular wastewater.

During the five days in which distinct sample peaks were obtained from the instrument, total carbon concentrations ranged from 50 to 65 percent of values measured using the laboratory organic carbon analyzer. Because of the more consistent operation of the laboratory instrument, it was apparent that values determined using the on-stream analyzer were incorrect. Laboratory operation of the total carbon analyzer prior to its installation in the continuous-flow system indicated that several injections (approximately 3 to 6) were necessary to obtain consistent readings from each sample. Additionally, it was found that the injection of 5 to 8 distilled water samples were necessary to purge the reaction chamber when changing from one sample to another. Since the multi-stream selector was designed to make only one injection of each waste sample alternated with one injection of distilled water, consistent operation of the instrument was not obtained.

During a two-month test period, operation of the total carbon analyzer was unsatisfactory. Even with a high degree of maintenance, the instrument was incapable of properly measuring and recording the total carbon of waste samples from various points in the treatment system. It is believed that problems encountered in using this instrument were due to the type of materials which are not uncommon in wastewaters produced in most diversified organic chemical plants.

SECTION IX

ACTIVATED SLUDGE INVESTIGATIONS

Activated sludge investigations were conducted in three phases. Phase I was initially conducted to determine the feasibility of activated sludge treatment for this waste. Phase II investigations were continued from the Phase I investigation to determine longer range performance of these units and to determine the effects of temperature on removal efficiencies. Phase III tests were conducted in order to compare differences in the response of the waste to treatment using air and oxygen aeration. The results of these investigations are described in this chapter.

AIR AERATION TESTS

During Phase I investigations, activated sludge units were fed raw equalized waste, unclarified trickling filter effluent, and clarified trickling filter effluent. The objectives of this portion of the investigation were to determine the removals of organic constituents possible using activated sludge treatment, to determine kinetic parameters required for process design, and to determine the effect of pretreatment in the present trickling filter system on the effectiveness of activated sludge treatment. Five 40-gal (151 liter) activated sludge units were operated during this portion of the study.

After a five-week stabilization period, the units were judged to be acclimated to the waste and the comprehensive testing program was begun. Data for these tests were collected over a seven-week period from January 11 to March 3, 1973.

Phase II investigations were initiated on March 10 and continued until May 26, 1973. One unit was placed outside and allowed to operate at the ambient temperature. Two other units were operated indoors at a temperature of approximately 18°C.

Substrate Removal

A summary of operating results for Phase I is shown in Table C1 of Appendix C. For units receiving the unclarified trickling filter effluent feed, the influent total BOD concentration averaged 2,270 mg/l. Total effluent BOD concentrations averaged 15, 53, and 233 mg/l, respectively, for units operated at F/M ratios of 0.08, 0.23, and 0.40. Percentage removal efficiency ranged from 92 to 99 percent. Thus, while the waste can be treated to low effluent BOD values, these values can be achieved only at low F/M ratios.

Table C2 contains data for all units operated at temperatures ranging from 14 to 18°C during Phase II. Table C3 contains data for a three-week period in which the temperature of the outside units averaged 6°C. Effluent BOD values were 26 and 71 mg/l, respectively, for units operated at 17 and 6°C, while the influent BOD was 1,100 mg/l.

The variation of soluble effluent BOD with the substrate removal rate for Phase I and Phase II is shown in Figure 31. These data show that the BOD removal rate constant, k, was 0.0045 1/mg-day. The average basin temperature during this study was 18°C.

From basic principles describing the activated sludge process, a linear relationship is expected between soluble effluent BOD and the substrate removal rate. Greater slopes from this relationship indicate greater amenability to activated sludge treatment. Deviation from this linear relationship is usually found at high substrate removal rates.

Because of the significant weekly variability of this waste, a statistical distribution of substrate removal rate constants was developed from data collected during this part of the investigation. On a BOD basis, the substrate removal rate constant, k, varied from approximately 0.0013 to 0.017 1/mg-day, with higher substrate removal rate constants concurrent with periods of clearer effluent. Because of the variability of manufacturing operations in the plant, the long-term observation of treatability characteristics of this waste is important for the proper design of an activated sludge system. For operation at constant F/M values, the effluent BOD will vary in proportion to the value of the substrate removal rate constant.

The principal effect of temperature changes on the activated sludge process is manifested in the effect on the substrate removal rate constant, k. A decrease in temperature will decrease the rate of biological metabolism which reduces the observed k value. For a system treating a given volume of waste, operation at reduced values of k will result in an increased BOD concentration. To determine the effects of low temperatures on the

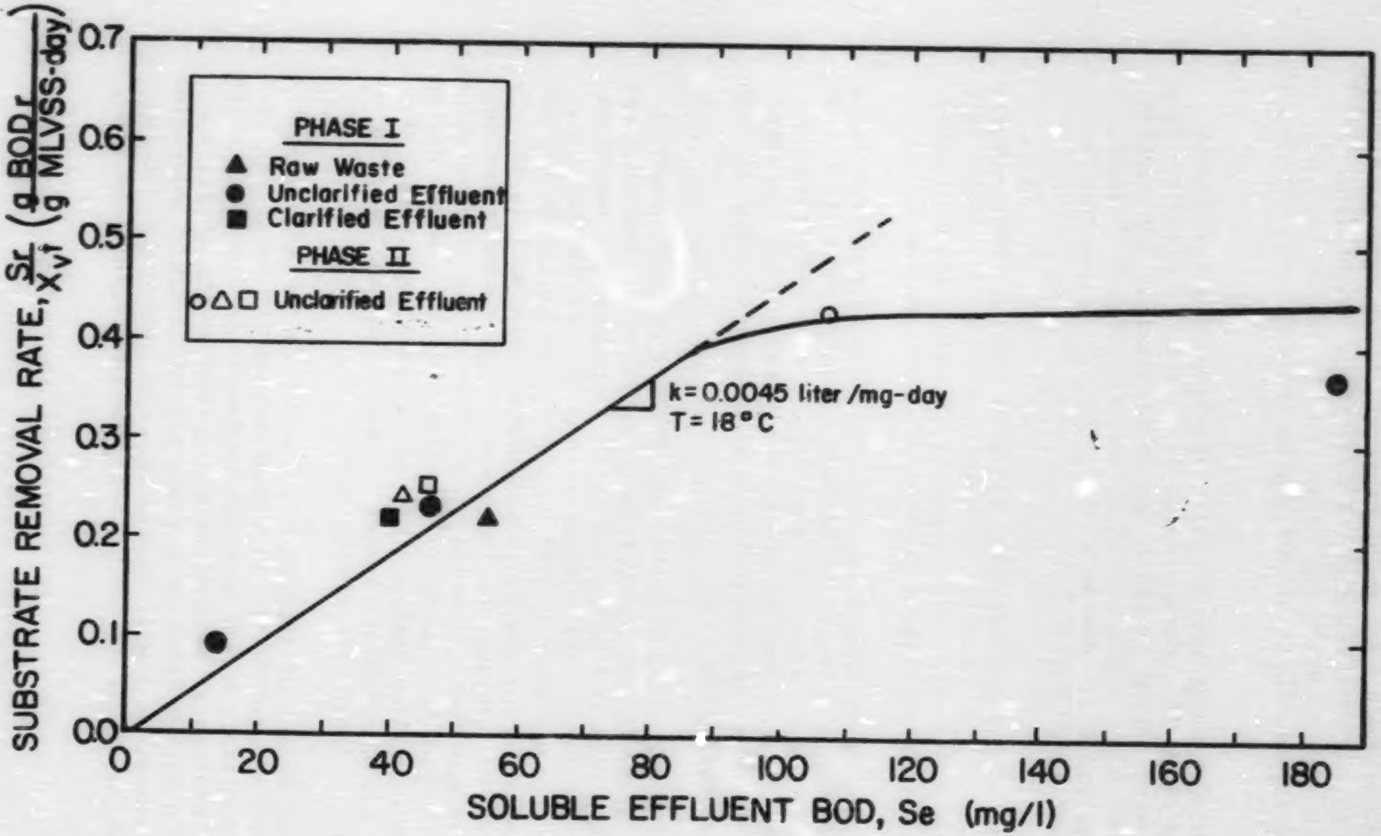


FIG. 31 . BOD REMOVAL RATE CONSTANT - PHASES I AND II biological process, all three units in Phase II were operated in parallel receiving the unclarified trickling filter effluent as feed. The relationship of soluble effluent BOD to the substrate removal rate at different temperatures is shown in Figure 32. The relationship used to relate the change of the substrate removal rate constant with temperature is expressed as:

$$k_{T_1} = k_{T_2}\theta^{T_1-T_2}$$

The symbol, θ, is the temperature coefficient for the system. From the BOD removal rate constants determined in Figure 32, the value of θ was determined to be 1.09. Due to the short sampling period the collection of additional data would be advisable to confirm this value.

Oxygen Requirements

Oxygen requirements may be calculated from a correlation of the substrate removal rate with the oxygen utilization rate expressed as grams oxygen utilized per gram MLVSS per day. The slope of the line defined by this relationship represents the assimilative oxygen utilization coefficient, a', while the intercept on the y-axis represents the endogenous oxygen utilization coefficient, b'.

The data illustrated in Figure 33 show that oxygen requirements for this waste are 1.12 g O_2/g BOD removed and 0.03 g O_2/g MLVSS-day. The oxygen utilization coefficient is somewhat higher than usually encountered for aerobic biological treatment systems, while the endogenous utilization coefficient is somewhat lower than usually measured.

Sludge Production

The production of biological sludge is directly proportional to the quantity of organic matter removed. The net sludge growth is the sum of the sludge growth resulting from the removal of organic constituents and the decrease in sludge mass due to endogenous respiration of the microorganisms.

Since sludge production is directly related to the organic loading rate, the net sludge production rate is correlated to the substrate

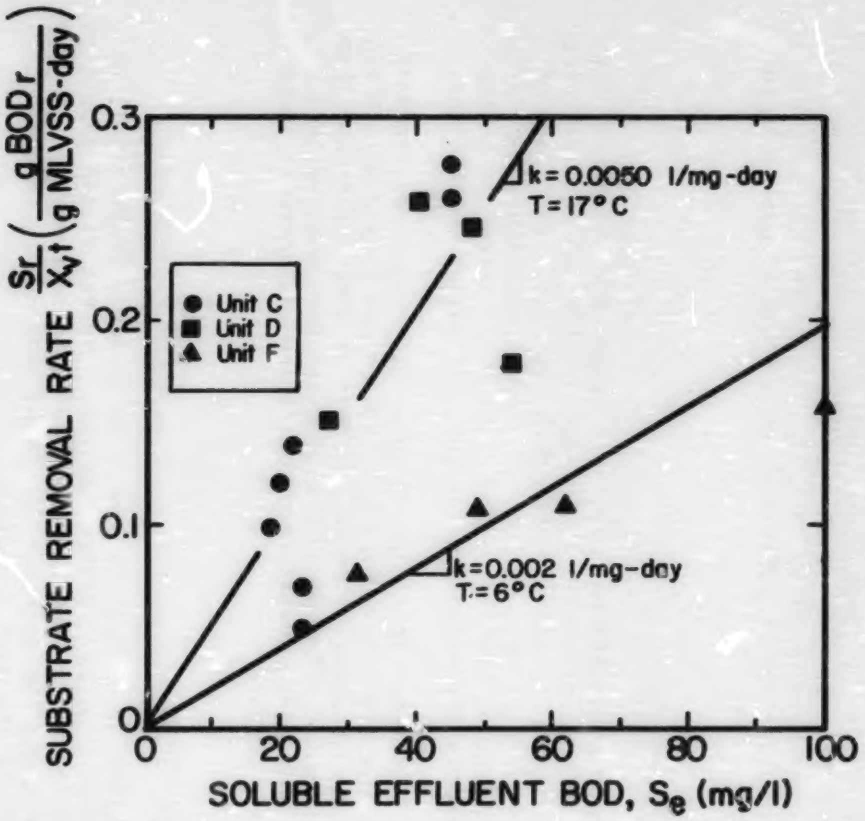


FIG.32. TEMPERATURE EFFECT ON ACTIVATED SLUDGE PERFORMANCE

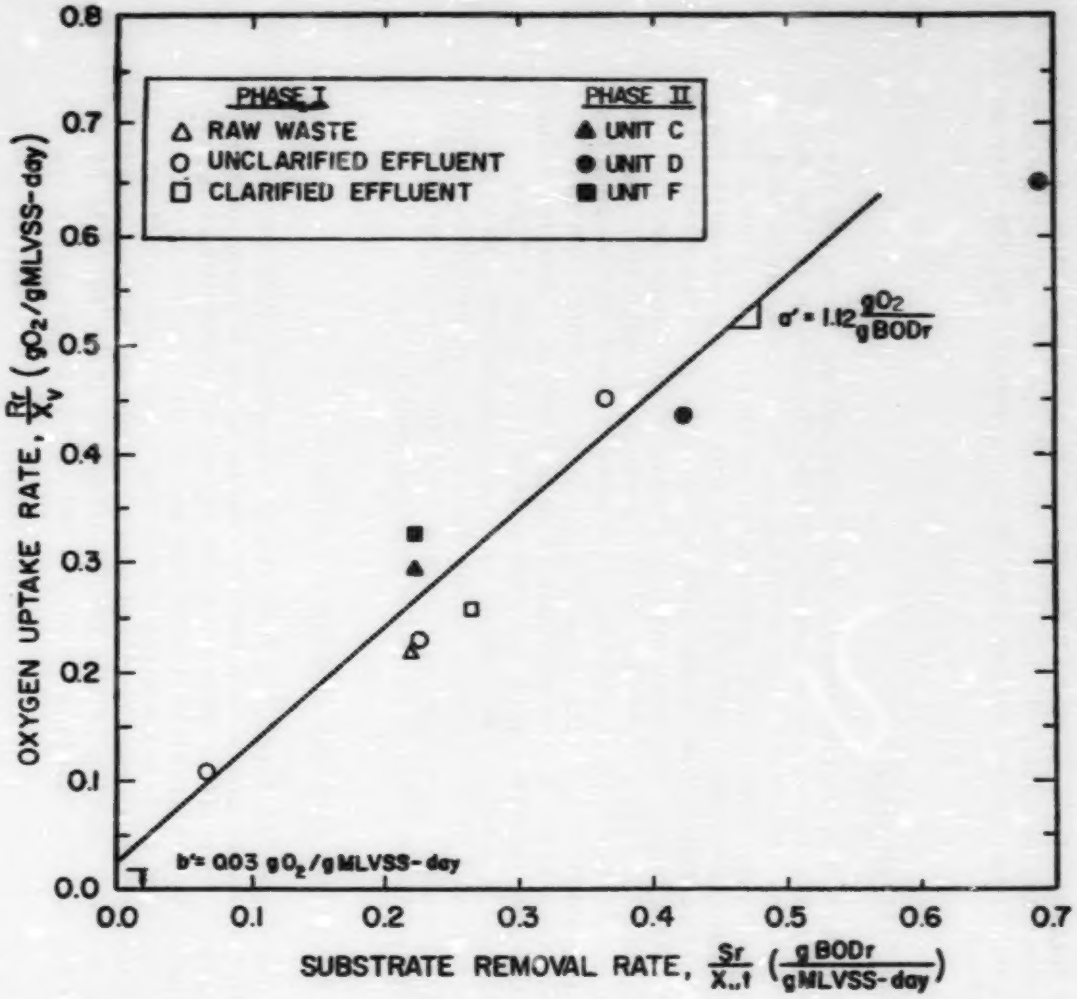


FIG. 33. DETERMINATION OF OXYGEN UTILIZATION COEFFICIENTS, BOD BASIS-PHASES I AND II

removal rate. The slope of the line represented from this relationship is the sludge yield coefficient, a, and the y-intercept is the endogenous decay rate, b. The data correlated in Figure 34 define a sludge yield coefficient of 0.37 g VSS/g BOD removed and an endogenous decay rate of 0.02 g VSS destroyed/g MLVSS-day.

In order to determine the total sludge solids produced, the volatile content of the wasted sludge must also be known. Measurements made over the duration of this study showed that the volatile fraction of mixed liquor solids ranged from 74 to 83 percent for the five units operated.

Sludge Settleability

Sludge settling, as measured by the sludge volume index and the zone velocity tests, was determined during the course of the investigation. Because sludge settling in small-scale units does not adequately simulate the performance which can be achieved in prototype clarifiers, additional tests were made to determine the probable concentration of effluent solids from full-scale activated sludge treatment. For most organic loadings the sludge volume index was less than 60 ml/g. While the data indicate that the best sludge settleability is obtained at higher organic loadings, effluent suspended solids data show that a higher concentration of solids is present in the effluent at high loadings. Average values for the zone settling velocity ranged from 13.6 to 24.1 ft/hr (4.15 - 7.34 m/hr). The highest zone settling velocity was obtained at an F/M of 0.23/day using the unclarified trickling filter feed. Compared to values usually obtained for these measures of sludge settleability, this sludge settled very well.

AIR AND OXYGEN AERATION COMPARATIVE TESTS

For the Phase III investigations two of the 55-gal (208 liter) activated sludge units were modified to be operated using oxygen aeration. In order to determine the effect of pretreatment by air stripping on the treatability of the waste, one air and one oxygen unit received a stripped waste feed. Two other units received a feed consisting of raw equalized waste. A summary of the operating data for this portion of the investigation is shown in Table B4 of Appendix B. The quantity of oxygen fed to the oxygen activated

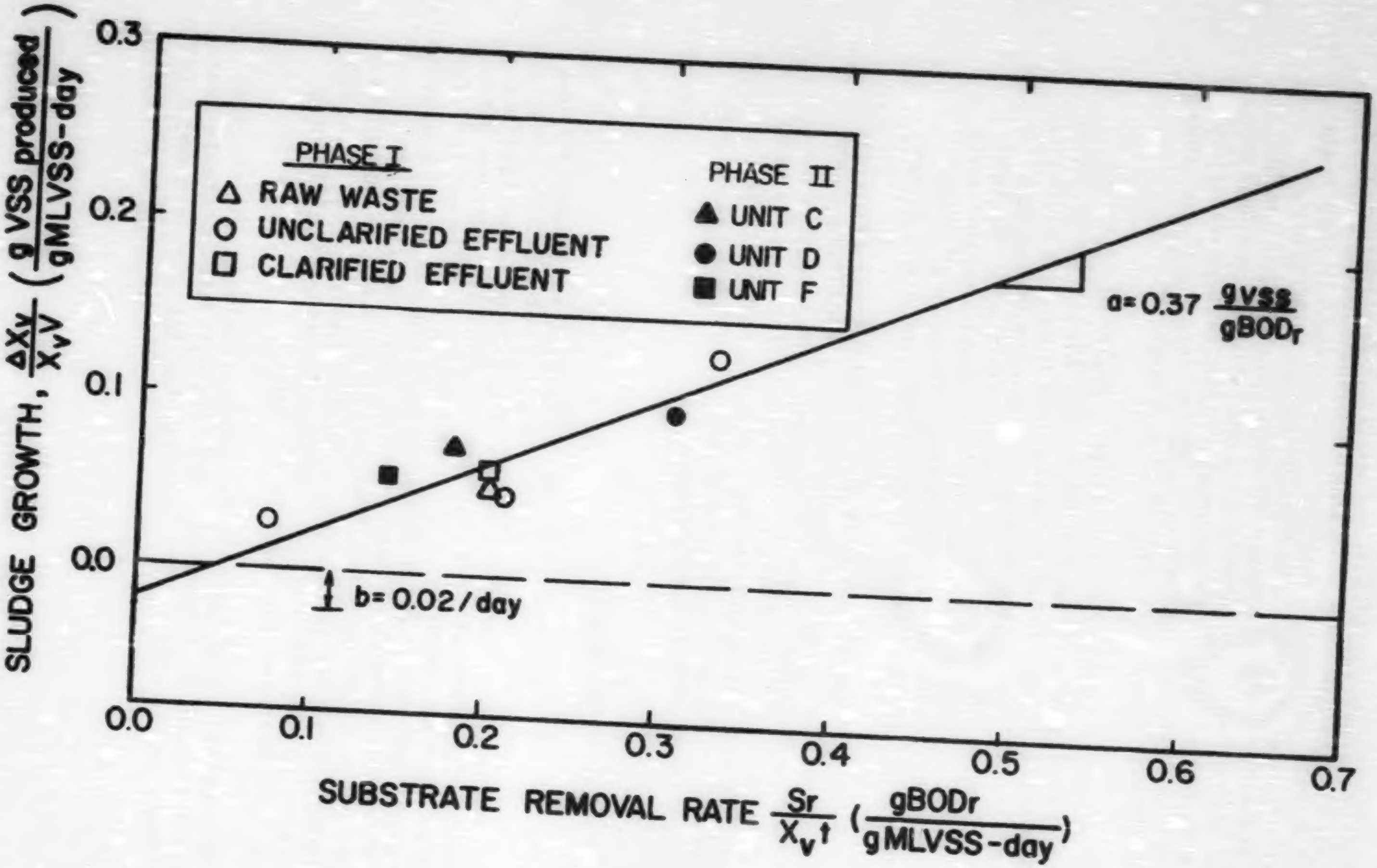


FIG. 34. EXCESS SLUDGE PRODUCTION, BOD BASIS - PHASES I AND II

sludge systems receiving unstripped and stripped waste required approximately 7 cu ft/day (0.2 cu m/day) to maintain a dissolved oxygen in the mixed liquor of 5 mg/l or higher and to maintain the oxygen content in the off-gas less than or equal to 50 percent. The average volumes of off-gas measured were greater than would be expected, indicating the presence of some leaks in the system. However, the volume of gas flowing through these systems was considerably less than the approximately 1,800 cu ft/day (51 cu m/day) which passed through the air activated sludge systems.

While oxygen aeration is usually technically feasible in cases where air activated sludge treatment may successfully be employed, several considerations warranted a feasibility study regarding the use of oxygen in this case. Because problems were encountered in Phase I tests due to the tendency of the aeration basin pH to decline below 6.0, there was concern that pH stabilization would prove even more difficult in oxygen systems containing a high partial pressure of CO_2 in the gas phase above the mixed liquor.

An additional indication of the need for a feasibility study was the apparent contribution of air stripping as a significant BOD removal mechanism in the air aerated systems. Because there was reason to believe that inhibitory constituents in the waste were being removed by air stripping, there was the possibility that these substances might accumulate in an oxygen aerated system and be detrimental to treatment performance.

Substrate Removal Relationship

The BOD removal rate constants for the unstripped waste are defined in Figure 35. The value of k for the air system was 0.0034 1/mg-day which was somewhat less than that found in previous investigations. This is more significant when it is realized that the average basin temperature for these tests was 23°C compared to a value of 18°C for previous investigations. The difference in these values is perhaps a result of the withdrawal of some cooling flow from the plant wastestream which might have resulted in the increase in concentrations of some inhibitory substances. The BOD removal rate constant for the oxygen system was 0.00083 1/mg-day, a value significantly lower than that measured for the air system.

Similar BOD removal relationships for units receiving the stripped waste feed are shown in Figure 36. BOD removal rate constants measured for these units were 0.0041 1/mg-day for the air unit and 0.0076

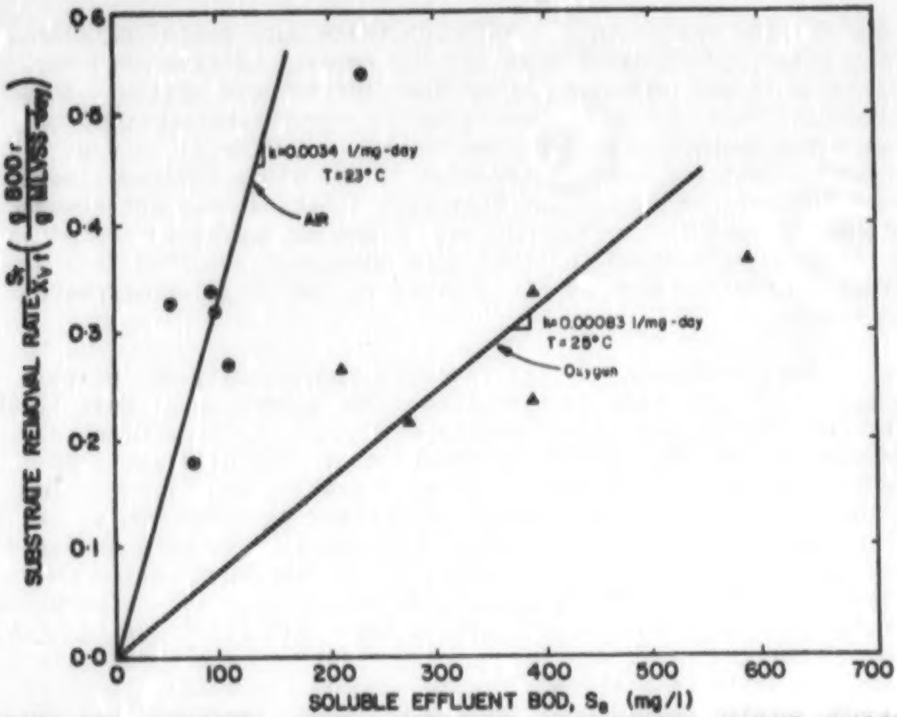


FIG.35. BOD REMOVAL RATE CONSTANT, UNSTRIPPED WASTE-PHASE III

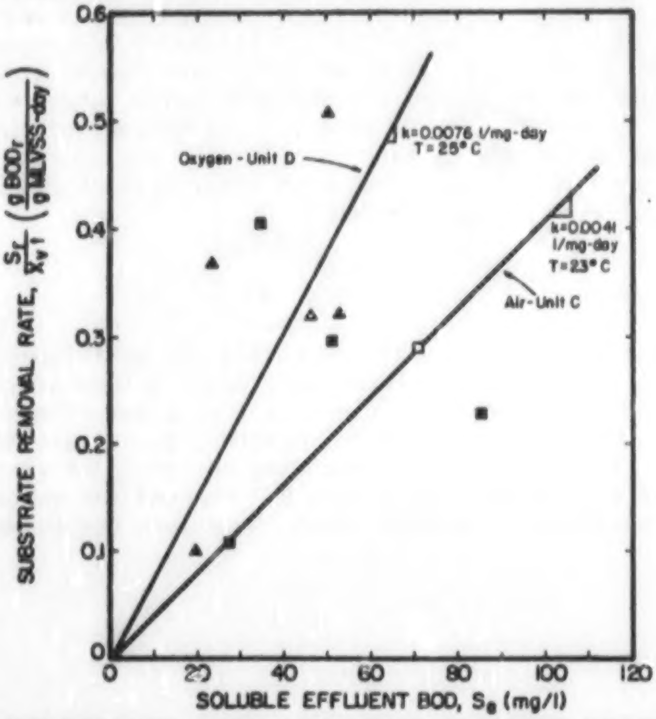


FIG. 36. BOD REMOVAL RATE CONSTANT, STRIPPED WASTE-PHASE III

1/mg-day for the oxygen unit. Although these data exhibited significant scatter, it appeared that the BOD removal relationship for the oxygen unit was somewhat better than for the air system. These data indicate that, in fact, some volatile constituent present in the raw wastestream was purged from the waste by the air stripping pretreatment and through aeration in the air activated sludge system. Further, it is evident that this substance was not removed, or not nearly to such a great degree, in the oxygen system receiving the unstripped waste feed and that this substance resulted in a significant deterioration of performance in the closed-atmosphere oxygen system.

Following the conclusion of this investigation on July 28, Units III-A and III-B were kept in operation. The operation of Unit III-B was changed from oxygen to air aeration in order to determine whether or not the BOD removal performance of this unit would approach that of Unit III-A under similar operating conditions. The chronological variations of control parameters for these units are shown in Figure 37. These data show that these units were operated under essentially identical conditions for a four-week period and that the effluent BOD and COD concentrations of Unit III-B decreased and became identical to effluent values from Unit III-A during this period. These data give further substantiation to the hypothesis that some volatile constituents accumulated in the mixed liquor of the oxygen-aerated units receiving the raw waste feed.

It is also significant to note that performance in the air system was somewhat improved following the air stripping pretreatment. This is evidenced by the k values of 0.0034 and 0.0041 1/mg-day for units receiving the unstripped and stripped waste feed, respectively. No adequate explanation for the superior performance of the oxygen units compared to the air system receiving the stripped waste feed can be offered which is consistent with other data observed for air and oxygen activated sludge treatment.

Oxygen Utilization

Oxygen utilization data for units receiving the unstripped waste feed are shown in Figure 38. From these data, values of the assimilative oxygen utilization coefficient, a', were determined to be 1.14 and 0.81 g O_2/g BOD removed for oxygen and air aeration, respectively. Values for units receiving the stripped waste feed (Figure 39) were 0.89 and 0.68 g O_2/g BOD removed for oxygen and air aeration, respectively. In both cases the values for oxygen units

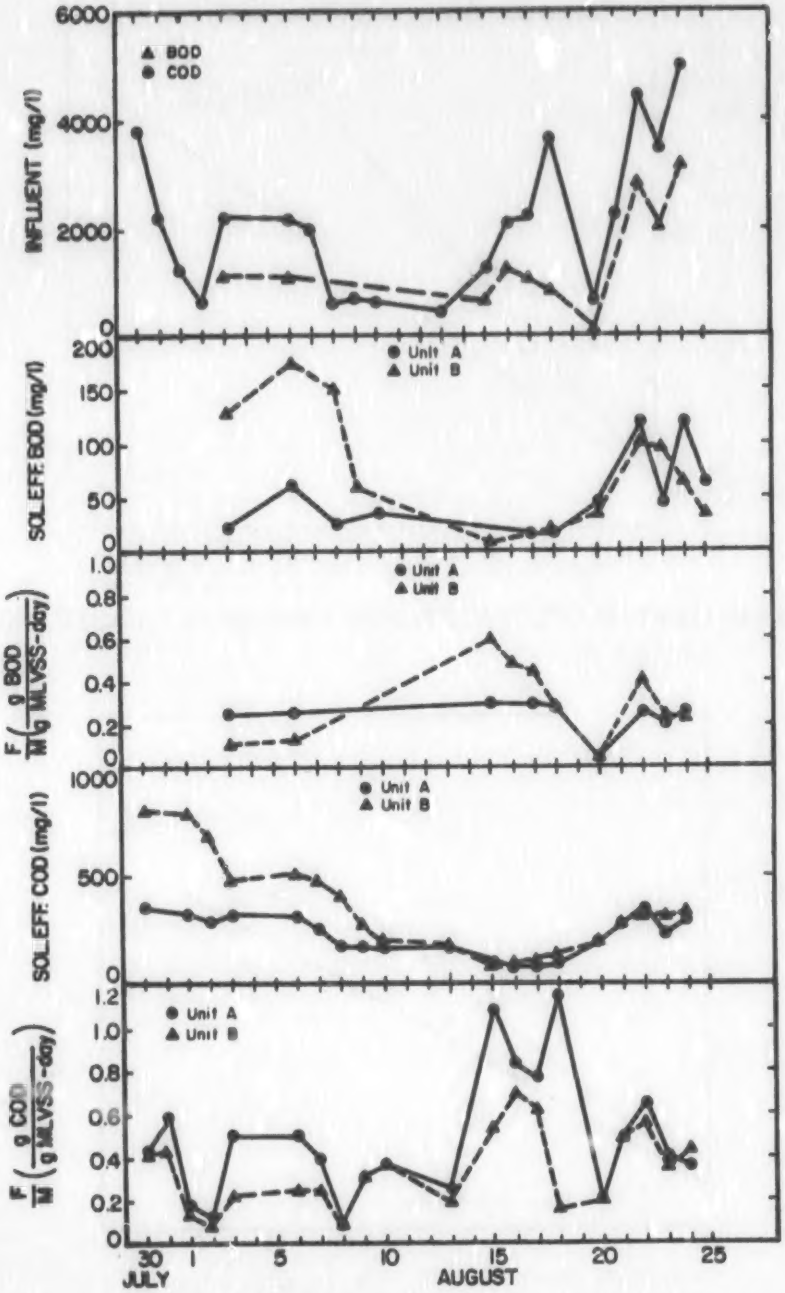


FIG. 37. PERFORMANCE OF OXYGEN-ACCLIMATED SLUDGE SUBJECTED TO AIR AERATION

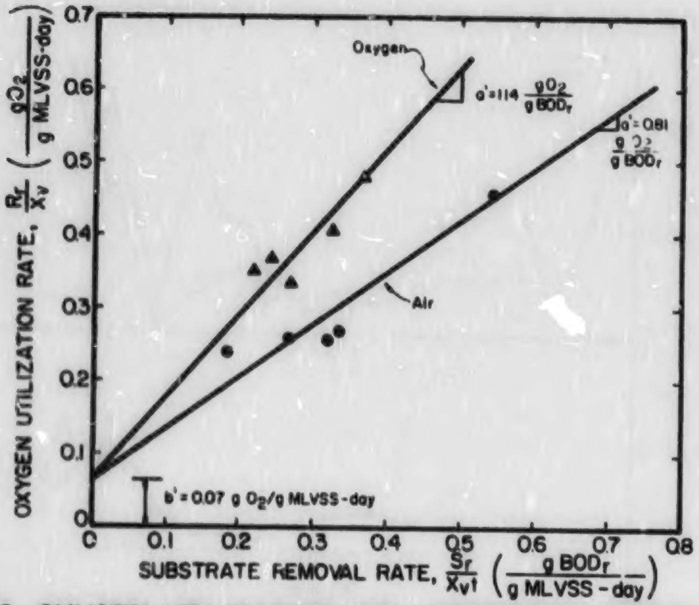


FIG.38. OXYGEN UTILIZATION FOR UNSTRIPPED WASTE, BOD BASIS - PHASE III

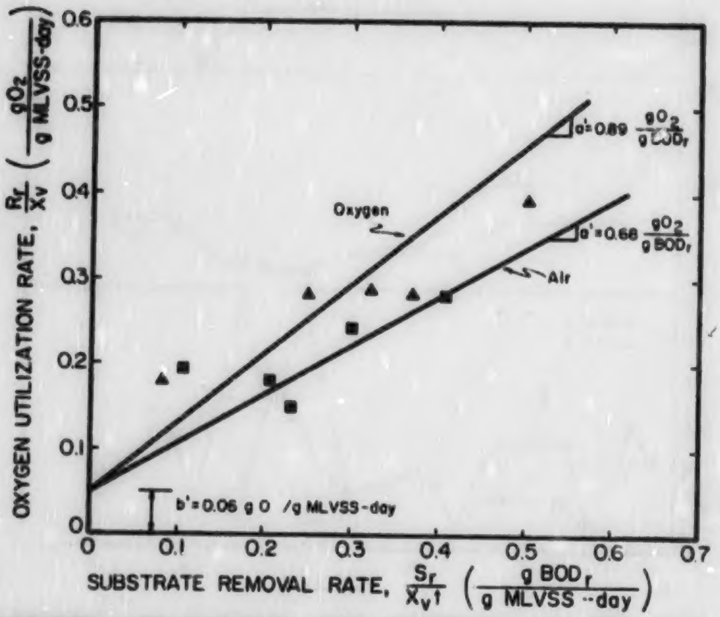


FIG.39. OXYGEN UTILIZATION FOR STRIPPED WASTE, BOD BASIS - PHASE III

were significantly higher than for air aeration indicating BOD removal by air stripping.

Sludge Settleability and Effluent Solids Concentration

Data summarized in Table B4 of Appendix B show that sludge settleability as measured by the sludge volume index and zone settling velocity tests was very good for all units. However, the units operated using oxygen aeration did exhibit lower SVI values of 37 and 48 ml/g (unstripped and stripped waste) compared to 55 and 61 ml/g for air aerated units. Values for the zone settling velocity were not as conclusive. While sludge settling measured for all units was very good, effluent suspended solids concentrations were high.

SUMMARY

Activated sludge investigations showed that the waste is amenable to treatment using this process. However, extremely low organic loadings (F/M ratios) were required to achieve effluent BOD values less than 30 mg/l. Process performance was significantly impaired using oxygen aeration in a closed atmosphere system treating the unstripped waste. Performance using oxygen aeration treating a stripped waste was equal to or better than that obtained with air aeration. However, the failure of the unstripped waste oxygen unit to produce acceptable BOD values raised questions regarding the feasibility of a staged-oxygen process treating a stripped waste.

Values for kinetic parameters determined in this investigation are presented in Table 9. Data correlated on a COD basis are also presented for comparison with BOD data. Some variation of data was apparent during the Phase III investigation when the plant was operating on a heavy production schedule. Sludge production data for Phase III were scattered; however, it appeared that the production was not significantly different from that observed in Phases I and II.

The BOD removal rate constants corrected to 20°C for different phases of the investigation are shown in Figure 40. Data points shown in this figure are averages for entire portions of the investigation. These data indicate the poor performance of oxygen

TABLE 9

SUMMARY OF ACTIVATED SLUDGE DESIGN PARAMETERS

Parameter	Phases I and II		Phase III			
			Unstripped Waste		Stripped Waste	
Aeration Temperature	Air 17	Air 18	Air 23	Oxygen 25	Air 23	Oxygen 25
BOD Basis						
BOD Removal Rate Constant[a], k (1/mg-day)	0.0048	0.0042	0.0034	0.00083	0.0041	0.0076
Sludge Yield Coefficient, a (g VSS produced/g BOD_r)	0.38	0.37	b	b	b	b
Endogenous Respiration Rate, b (g VSS destroyed/day-g MLVSS)	0.02	0.02	c	c	c	c
Assimilative Oxygen Utilization Coefficient, a' (g O_2/g BOD_r)	1.11	1.14	0.81	1.14	0.68	0.89
Endogenous Oxygen Utilization Coefficient, b' (g 02/g MLVSS-day)	0.03	0.02	0.07	0.07	0.06	0.06
COD Basis						
COD Removal Rate Constant, k (1/mg-day)	0.0025	0.0024	0.0016	0.00097	0.0048	0.010
Sludge Yield Coefficient, a (g VSS produced/g COD_r)	0.18	0.25	d	d	d	d
Endogenous Respiration Rate, b (g VSS destroyed/day-g MLVSS)	0.02	0.02	e	e	e	e
Assimilative Oxygen Utilization Coefficient, a' (g O_2/g COD_r)	0.67	0.79	0.62	0.78	0.43	0.57
Endogenous Oxygen Utilization Coefficient, b' (g O_2/g MLVSS-day)	0.03	0.02	0.055	0.055	0.055	0.055

[a] Values given for temperatures at which investigations were made.

[b] Average value of the sludge yield coefficient was 0.30 g VSS produced/g BOD_r. However, data were not sufficiently accurate to permit determination of values for each unit.

[c] Accurate determination of b could not be made from data available. However, value was not significantly different from 0.02/day determined in earlier tests.

[d] Average value for all units was 0.18 g VSS produced/g COD_r.

[e] Average value appeared to be 0.02/day.

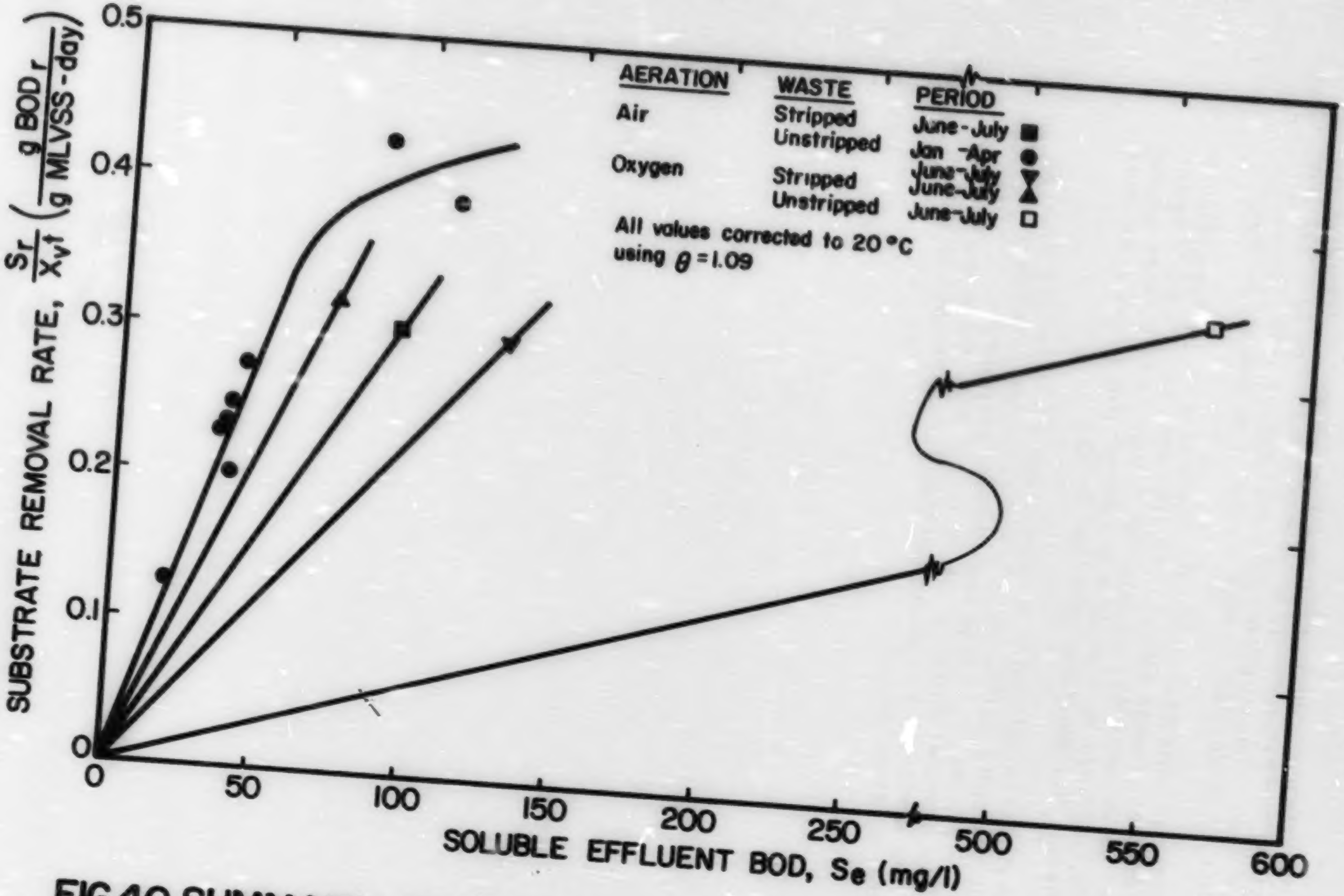


FIG.40.SUMMARY OF BOD REMOVAL RELATIONSHIPS FOR ACTIVATED SLUDGE SYSTEMS

aeration for the unstripped waste and the beneficial effect which air stripping pretreatment had on the performance of both air and oxygen aeration. The great weekly variations of k for Phases I and II must be taken into account in the final process design.

Oxygen transfer characteristics for the activated sludge effluent were determined several times during the investigation. The minimun alpha value was 0.45. Repetition of the tests in warm weather generally confirmed earlier data and demonstrated that a value of approximately 0.4 should be used for design.

In general, activated sludge treatment was characterized by high effluent suspended solids (90 - 175 mg/l), high refractory effluent COD (275 - 500 mg/l), a high temperature dependence ($\theta = 1.09$), low oxygen transfer characteristics ($\alpha = 0.45$), and a low BOD removal rate constant, k. Sludge settleability was good throughout most of the investigation.

SECTION X

ACTIVATED CARBON INVESTIGATIONS

Activated carbon investigations were performed to determine the feasibility of this process for achieving additional removal of organic substances from the biological effluent. Initially, batch isotherm tests were conducted to determine the amenability of various carbons for use in treating this waste and to determine the effect of pH on adsorption efficiency. Subsequent tests were performed in column systems to determine breakthrough characteristics for the most promising carbons identified in the batch tests. While most column tests were conducted using activated sludge effluent to determine removals of refractory organic substances, some column tests were conducted using the more concentrated trickling filter effluent. Results of both batch and column tests are summarized in this section.

ISOTHERM TESTS

Batch isotherm tests were conducted at room temperature (approximately 20°C) by contacting the carbon with activated sludge effluent for a 24-hr period in flasks attached to a wrist-action shaker. In most cases, several runs were made for each individual carbon in order to minimize the effect of variations in wastewater composition. It was found that while the adsorption capacities obtained for individual runs might vary significantly with wastewater composition, the relative ranking of various carbons by capacity remained basically unchanged for most of the runs.

Results from the isotherm tests for COD removal were correlated using the Freundlich isotherm. However, color removal data did not follow this isotherm and was correlated using the relationship:

$$y = A\ (X^k)$$

where y is the equilibrium in color concentration in APHA units, X is the carbon dose in mg/l, A is the ordinate intercept in APHA units, and k is a constant determined from the slope of the line obtained plotting equilibrium color concentration versus carbon dose. Using this relationship lower values of A and increasing values of the absolute k indicate increasing affinity for color-producing substances.

Carbon isotherms were measured with Hydrodarco 4,000, Filtrasorb 400, and Westvaco Aqua Nuchar 8 x 30 carbons. Although nine different carbons were tested, these carbons showed the best performance for COD removal in initial isotherm screening tests, color removal, and combined COD and color removal characteristics. A summary of the data obtained from these isotherms is presented in Table 10. Calgon Filtrasorb 400 and Westvaco 8 x 30 showed the highest affinity for COD removal, while the Hydrodarco and Westvaco carbons showed the greatest affinity for color removal.

TABLE 10

CARBON ISOTHERM RESULTS

USING ACTIVATED SLUDGE EFFLUENT

Carbon	COD isotherm		Color isotherm	
	extrapolated capacity, X/M (g COD/ g carbon)	$\frac{1}{n}$	k APHA units (mg/l)	A (APHA units)
Hydrodarco 4000	0.6	0.86	0.650	67
Filtrasorb 400	1.3	0.50	0.264	220
Westvaco aqua nuchar 8 x 30	4.8	1.80	0.585	120

Additional batch tests were made to determine the effect of pH on adsorption performance. Tests were run as described previously except that the pH of the wastewater sample was adjusted prior to contacting with the carbon. Following the 24-hr equilibration period, the sample was filtered and the pH adjusted back to 7.0 prior to analysis for residual COD and color. The results of these tests are presented in Figure 41. Because it was necessary

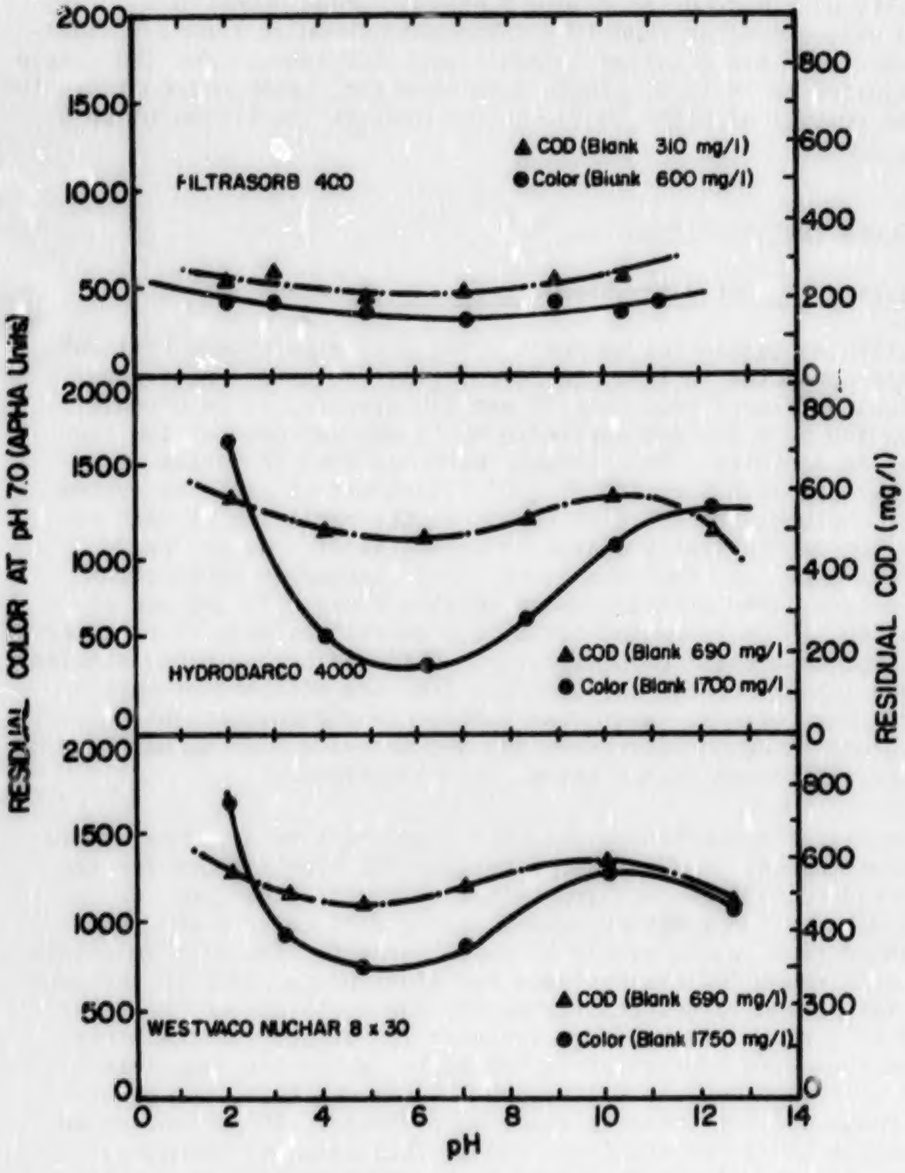


FIG. 41. INFLUENCE OF pH ON CARBON ADSORPTION CAPACITY

to run these tests on different days, the influent COD and color concentrations were different for the tests with Filtrasorb 400 than with the other two carbons. Some precipitation occurred in tests with Hydrodarco 4000 and Westvaco Aqua Nuchar 8 x 30 at pH values greater than 10.0. The data shown in Figure 41 indicate that this resulted in additional COD removal for the sample adjusted to pH 12.0. These data show that optimum conditions for the removal of both COD and color occur in the region of pH 4 to 7.

COLUMN INVESTIGATIONS

Tests Using Activated Sludge Effluent

Activated carbon column tests using activated sludge effluent were conducted in 1-in. (2.54 cm) glass columns. Activated sludge effluent from Unit IE and IID operated at an organic loading of 0.4 g BOD applied/g MLVSS-day was used as the feed for these tests. Four columns were operated in series at a flow rate of 2.5 gpm/sq ft (102 liters/min-sq m). The system was charged with 170 g of carbon at the beginning of each run. Carbon was initially contacted with distilled water for 24 hr to expel air from the pores, then charged to each column and backwashed with tap water until a clear effluent was obtained. High suspended solids concentrations made it necessary to backwash these columns daily. Prior to backwashing, samples were taken from the influent and from the effluent of each of the four columns. Waste was applied to the columns upflow. Hydrodarco 4000, Filtrasorb 400 and Westvaco Aqua Nuchar 8 x 30 were all tested in the column investigations.

Normalized breakthrough data for these runs on a TSC basis are shown in Figures 42 - 44. Immediate TSC breakthrough for the Hydrodarco 400 run was approximately 15 to 20 percent, while values for the Westvaco Aqua Nuchar and Filtrasorb 400 runs ranged from approximately 30 to 40 percent. The high immediate breakthrough for the Westvaco run might be due to high influent total carbon values (approximately 230 mg/l) during the first of this run, while average influent TSC values for the other two runs were approximately 120 mg/l. While the immediate breakthrough for the Filtrasorb 400 run was high, the breakthrough did not increase above approximately 40 percent on an average basis for the first 800 BV (bed volumes) throughput. However, as will be discussed in the next section, several cycles of regeneration in actual use might reduce this initial advantage.

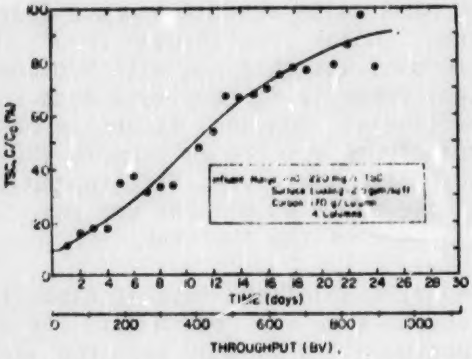


FIG.42.BREAKTHROUGH CURVE - HYDRODARCO 4000 CARBON
(cum/min-sqm=gpm/sqft x.004)

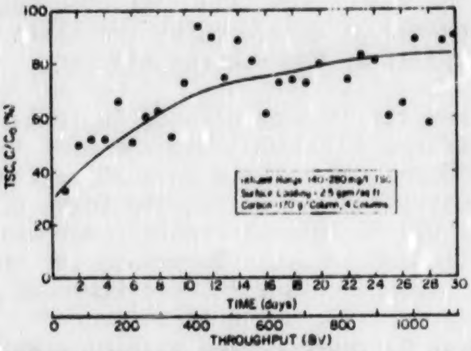


FIG. 43. BREAKTHROUGH CURVE - WESTVACO 8 × 30 CARBON
(cum/min-sqm=gpm/sqft x.004)

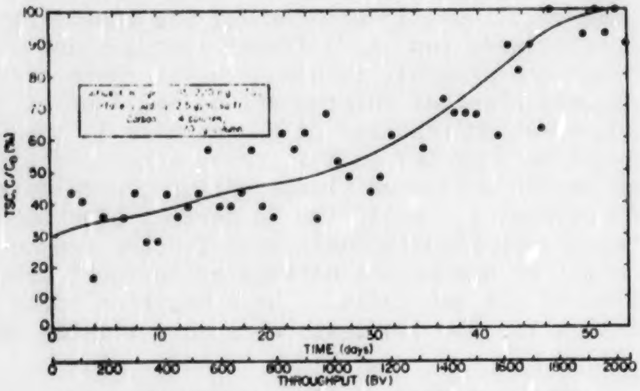


FIG. 44. BREAKTHROUGH CURVE - FILTRASORB 400 CARBON
(cum/min-sqm=gpm/sqft x.004)

Normalized color breakthrough curves for these same tests are shown in Figures 45 and 46. Although no entirely acceptable routine method exists for measuring color in industrial wastes, measurements were made using APHA units since the waste color closely matched these standards. Color breakthrough relationships show that the best color removal was obtained with Hydrodarco 4000 with progressively less capacity for Westvaco Aqua Nuchar and Filtrasorb 400, respectively. However, it should be pointed out that while color breakthrough for the Filtrasorb 400 carbon rose above 30 percent after approximately 75 BV throughput, the color breakthrough did not rise above 50 percent for any significant period during the remainder of the test run.

In order to determine any significant loss of capacity on regeneration, the spent Hydrodarco 4000 was returned to the manufacturer for regeneration. Subsequently, another test run was made in order to compare the performance of the carbon before and after one regeneration cycle. The TSC breakthrough curve for the regenerated carbon is shown in Figure 47. For this test, the adsorption kinetics were very favorable as evidenced by the sharp breakthrough approximately nine days after the run was started.

A comparison of the results is presented in Table 11. These results show that Hydrodarco 4000 exhibited the best removal characteristics for both COD and color based on a 30 percent breakthrough of the influent concentration. However, the total COD exhaustion capacity to approximately 90 percent breakthrough was greater for the Westvaco and Filtrasorb carbon. Reference to Figure 44 shows that while Filtrasorb 400 reached a TSC breakthrough of approximately 40 percent very quickly, the breakthrough did not increase above 40 percent for an extended period of time. While the 30 percent breakthrough capacity for COD was almost identical for the virgin and regenerated Hydrodarco 4000 carbon, the 90 percent COD breakthrough was significantly greater for the regenerated carbon. Because these tests were run at different times, changes in wastewater composition are probably responsible for these differences. However, it is possible that enlargement of small pores in the carbon upon regeneration resulted in an increase in the COD capacity. This might be expected in this case, since organic constituents contained in the activated sludge effluent were quite probably very large molecules. While the 30 percent breakthrough capacity for color was significantly less for the regenerated carbon, this might be due to the particular influent color variation at the time of the two tests. The exhaustion color breakthrough capacities for the two tests were more similar in value.

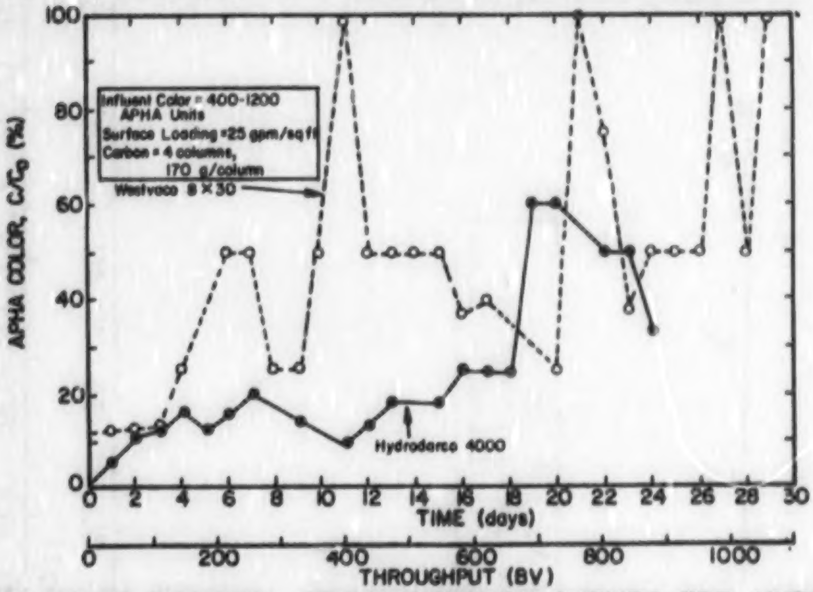


FIG. 45. COLOR REMOVAL BREAKTHROUGH CURVES FOR HYDRODARCO 4000 AND WESTVACO 8 × 30
(cu m/min-sq m = gpm/sq ft x 0.04)

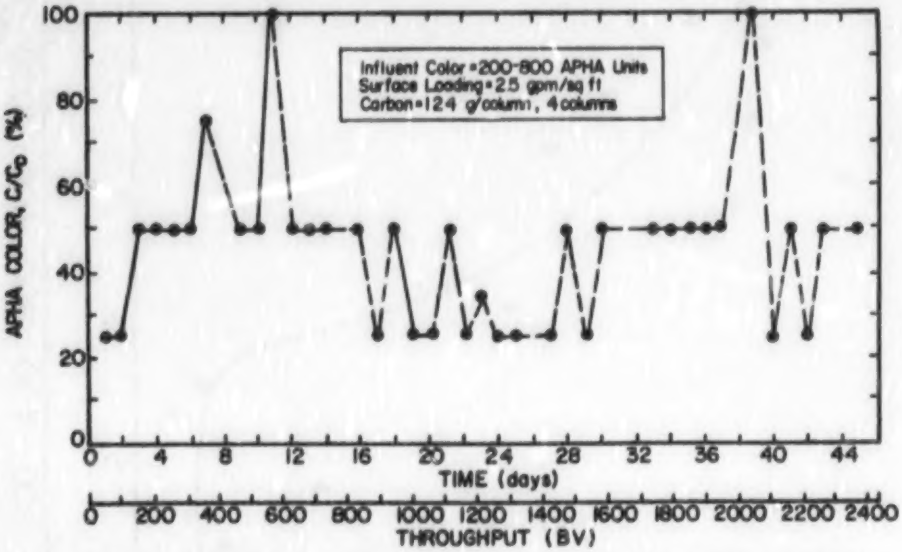


FIG. 46. COLOR REMOVAL BREAKTHROUGH CURVE FOR FILTRASORB 400
(cu m/min-sq m = gpm/sq ft x 0.04)

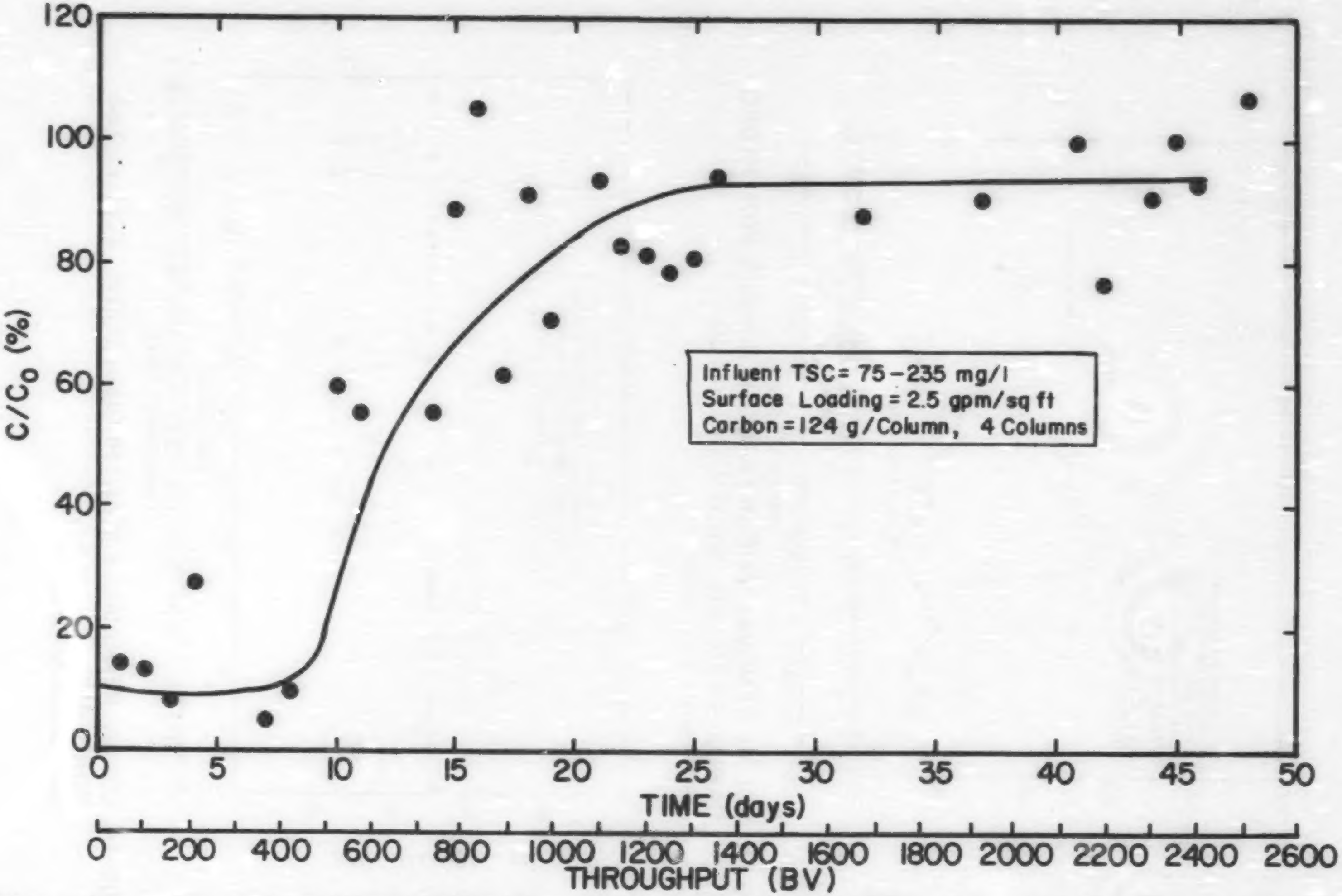


FIG. 47. TSC BREAKTHROUGH FOR REGENERATED HYDRODARCO 4000
(cu m/min-sqm = gpm/sq ft x 0.04)

TABLE 11

COMPARISON OF RESULTS FOR ACTIVATED CARBON ADSORPTION OF ACTIVATED SLUDGE EFFLUENT

Parameter	Hydrodarco 4,000		Westvaco 8 x 30	Filtrasorb 400
	Virgin	Regenerated		
30% Breakthrough capacity (g COD/g carbon)	0.27	0.26	0[a]	0.01
90% Breakthrough capacity (g COD/g carbon)	0.44	0.60	0.60	0.85
30% Color breakthrough capacity $\frac{\text{(APHA units x liters throughput)}}{\text{(g carbon)(1000)}}$	1.12	0.40	0.34	0.13
Exhaustion color breakthrough capacity $\frac{\text{(APHA units x liters throughput)}}{\text{(g carbon)(1000)}}$	1.39	1.25	1.58	0.95
Color breakthrough at end of run (%)	50	75	100	100

[a] Immediate breakthrough to greater than 30 percent.

Column Tests Using Trickling Filter Effluent

Additional column tests were conducted in 4-in. (10 cm) carbon columns to examine treatment of the trickling filter effluent using activated carbon. The results of these tests are shown in Figure 48. While a longer contact time might have resulted in additional COD removal, these results might be expected since the trickling filter effluent contained considerable concentrations of undegraded low molecular weight solvents which are not readily sorbable on activated carbon.

Greater degrees of color removal were achieved in these tests showing the greater sorbability of color molecules compared to other organic substances in the waste. Based on color analyses using the APHA standards, the Filtrasorb and Hydrodarco 3000 carbons averaged approximately 50 percent color removal throughout most of this run. Color removal through the Hydrodarco 3000 was more erratic, though. These tests indicate that further reductions of gross organics would be necessary in the trickling filter system before this combination of treatment processes would accomplish much COD removal. However, if color removal were a major treatment objective, such an approach would be more attractive.

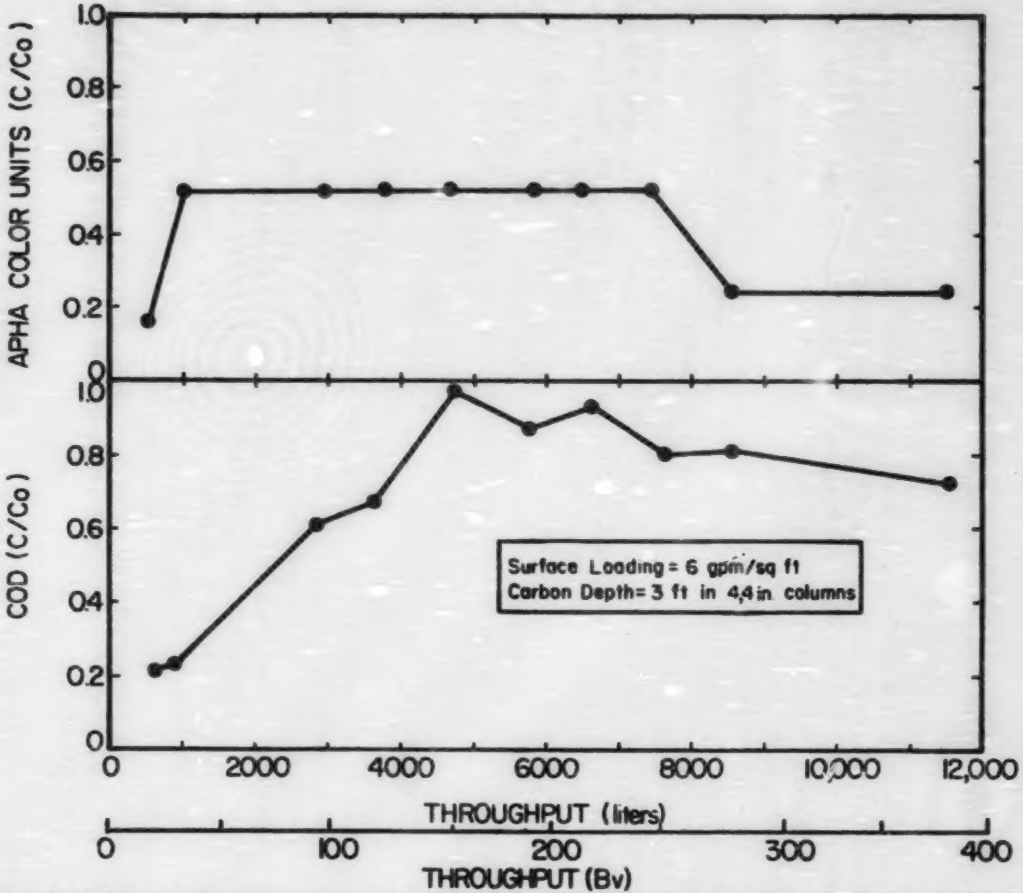


FIG.48.COD BREAKTHROUGH FOR FILTRASORB 400 USING TRICKLING FILTER EFFLUENT
(cu m/min - sq m = gpm /sq ft x 0.04)

SECTION XI

REMOVAL OF ORGANIC CONSTITUENTS BY AIR STRIPPING

Due to the large proportion of organic solvents in this wastestream, the removal of organic substances by air stripping was investigated to determine both the incidental removal by this mechanism in biological treatment processes and removals which could be obtained using air stripping as a unit treatment process. Preliminary laboratory experiments in which raw waste was stripped using diffused air aeration showed that approximately 75 percent removal of COD could be achieved in three days and approximately 92 percent could be removed in ten days. Because no biological inhibitory agent was added to the waste during these tests, it is probable that some removal of organic constituents was achieved by biological mechanisms. However, it seemed unlikely that the major part of the organics could be removed biologically during this period in the unseeded raw waste. Based on these results, additional tests were undertaken both in the laboratory and on a pilot-scale.

BATCH LABORATORY-SCALE TESTS

Initially, batch stripping tests were performed in the laboratory to determine the degree of organics removal which could be achieved by diffused air stripping and to determine the effect of temperature on these removals. Tests were performed in 2-liter graduated cylinders initially containing 1,000 ml of raw equalized waste. The laboratory compressed air was passed through an oil trap and diffused through the sample using a porous stone. Air flows were measured using a rotometer. In order to measure the effect of temperature on organics removal, the 2-liter cylinder was partially immersed in a constant-temperature bath and the temperature adjusted using ice and hot water.

The results of batch experiments run at temperatures of 5, 17, and 38°C are shown in Figure 49. While first-order removal of organic constituents would be anticipated for individual organic constituents, the removal of organic carbon in a multi-component system would be the summation of removals for individual organic constituents occurring at different rates. The semi-logarithmic plot of the data in Figure 49 corresponds to these mechanisms, although it does not confirm the hypothesis. Although all tests were not run for an extended period of time, it appears from available data that

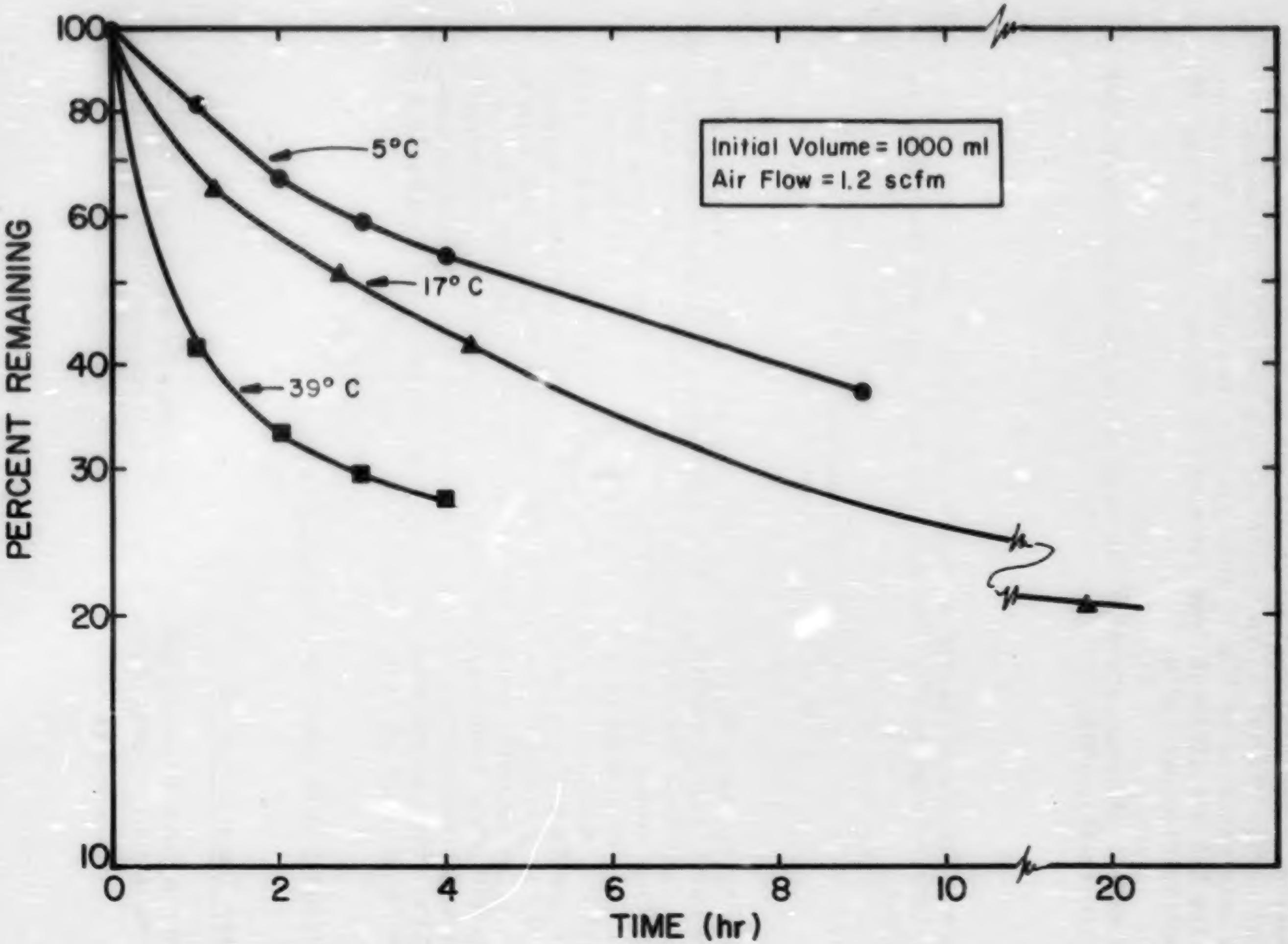


FIG. 49. TSC REMOVAL IN BATCH STRIPPING TESTS
(std. cu m/min = s cfm x 0.028)

the residual organic concentration at all temperatures tended toward a common value indicating the presence of a non-strippable portion of organics in the waste. The greatest removal of total carbon was achieved in the 17°C test which was run over a 20-hr period. The total carbon removal in this case was approximately 80 percent. The determination of BOD in this testing indicated that 96 percent of the BOD was stripped from the waste in 20 hr. The residual BOD concentration was 130 mg/l.

The effect of temperature on the rate of stripping was correlated to the relationship:

$$(k_s)_T = (k_s)_{T'} \ \theta^{T-T'}$$

where k_s is the stripping waste constant in hr^{-1} and T and T' are temperatures in °C. The rate constant, k_s, was calculated as follows:

$$k_s = \frac{\%\ \text{TSC remaining at } t_1}{t_1}$$

where t_1 is the time from the beginning of the test in hours. The data correlated according to this relationship is shown in Figure 50. These data show that the temperature coefficient, θ, decreased with time from a value of 1.043 for the first hour of these tests to 1.023 for a 4-hr period. The value during the first hour of 1.043 was almost identical to the value of 1.044 derived from data obtained from stripping tests performed with 1-component systems of acetone, alcohol, and methyl ethyl keytone found in the literature. The decrease of the temperature coefficient with time observed in this investigation would be expected for data obtained from batch experiments in which the residual total carbon concentration after an extended aeration time would be approximately the same for a range of test temperatures.

CONTINUOUS-FLOW STRIPPING INVESTIGATIONS

Packed Tower Stripping

Stripping was investigated in a packed tower by adding a 2,500 cfm (71 cu m/min) blower to the first pilot trickling filter. Tests

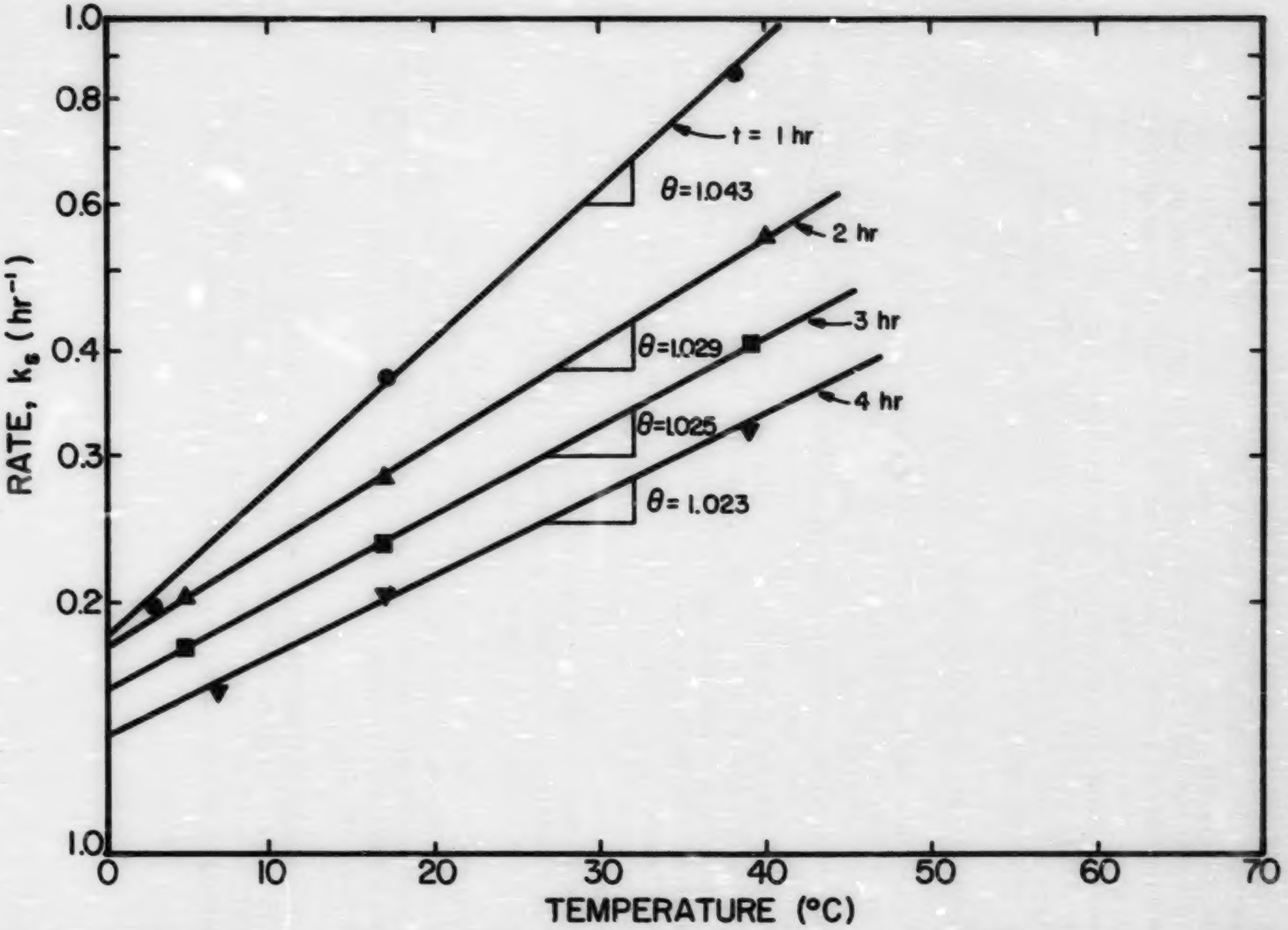


FIG.50.VARIATION OF STRIPPING REMOVAL COEFFICIENT WITH TEMPERATURE

were conducted at surface loadings varying from 1 - 10 gpm/sq ft (41 - 410 l/min-sq m) and air flows ranging from 300 - 2,500 cfm (8.5 - 71 cu m/min). Results obtained from this investigation on a BOD basis are shown as a function of the gas-liquid ratio in Figure 51. The highest BOD removal observed in these tests was approximately 60 percent and occurred most consistently at a gas-liquid ratio of 500 cu ft/gal (3.75 cu m/liter). Assuming that the stripping action in a continuous-flow system may be approximated by a first-order relationship, the data from these investigations is shown in linearized form in Figures 52 and 53. The mathematical relationships defining these data may be used to estimate BOD and TSC removals for full-scale systems having a depth of 20 ft (6.1 m).

Stripping in an Aerated Basin

Additional air stripping investigations were conducted in a 24-ft (7.3 m) diameter plastic swimming pool containing a 1.5 hp surface aerator. The volume of the basin was approximately 13,500 gal (51,000 liters) and the applied power level corresponded to approximately 150 hp/mil gal (29.5 kw/mil liter). Tests were conducted at detention times of 1.5 and 3 days. Results of these tests for the 1.5-day detention time on a chronological basis are shown in Figure 54. Total carbon removals of this period averaged approximately 40 percent. As an indication of biological activity which occurred in this basin, oxygen uptake tests for the influent and effluent of the basin were made periodically. The results of these tests shown in Figure 54 indicate significant biological activity in the waste. From these data it was estimated that approximately 30 percent of the BOD removed in this basin was due to biological mechanisms. A summary of results is shown in Table 12.

Chronological data for a 3-day detention time are shown in Figure 55. Because of the relatively high oxygen uptake rates observed in the basin, an additional test was run using the 3-day detention time in which 20 mg/l Cu was added to the basin influent in order to inhibit biological growth.

Only one oxygen uptake measurement was made during the period June 15 - 30. This measurement, made near the end of this run on June 28, was 52 mg/l-hr. Oxygen uptake measurements made during the period in which copper was added to the basin were 1.0, 6.0,

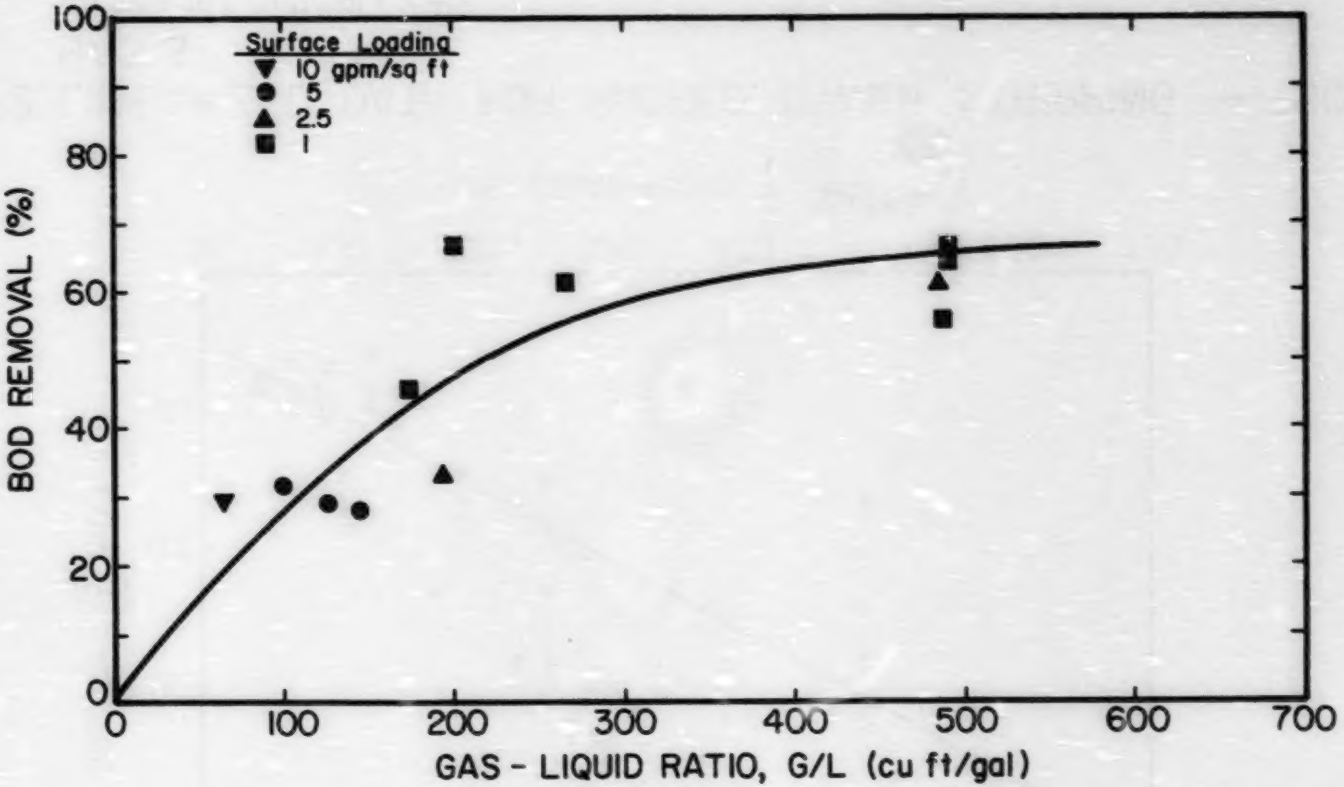


FIG.51 . PACKED TOWER AIR STRIPPING RESULTS, BOD BASIS
(cu m/min - sq m = gpm/sq ft x 0.04; cu m/cu m = cu ft/gal x 7.48)

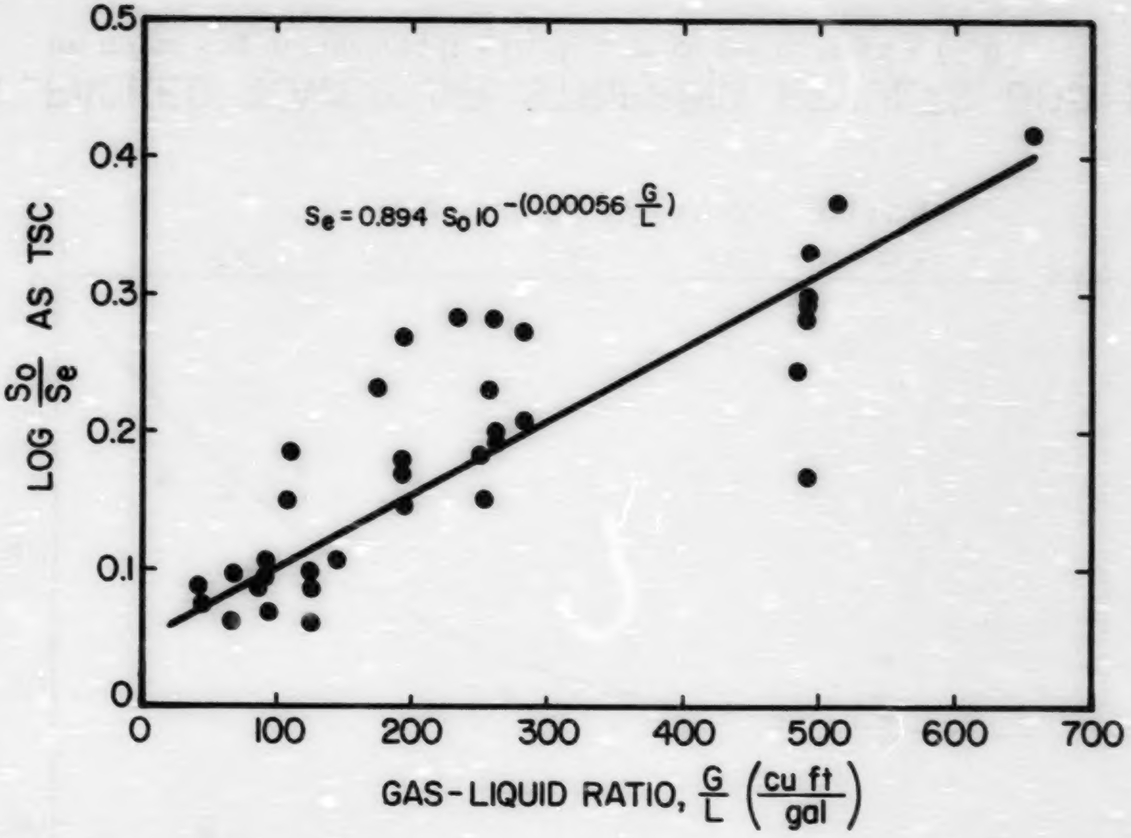


FIG.52. LINEARIZED DATA FOR PACKED TOWER STRIPPING — TSC BASIS
(cu m/cu m = cu ft/gal x 7.48)

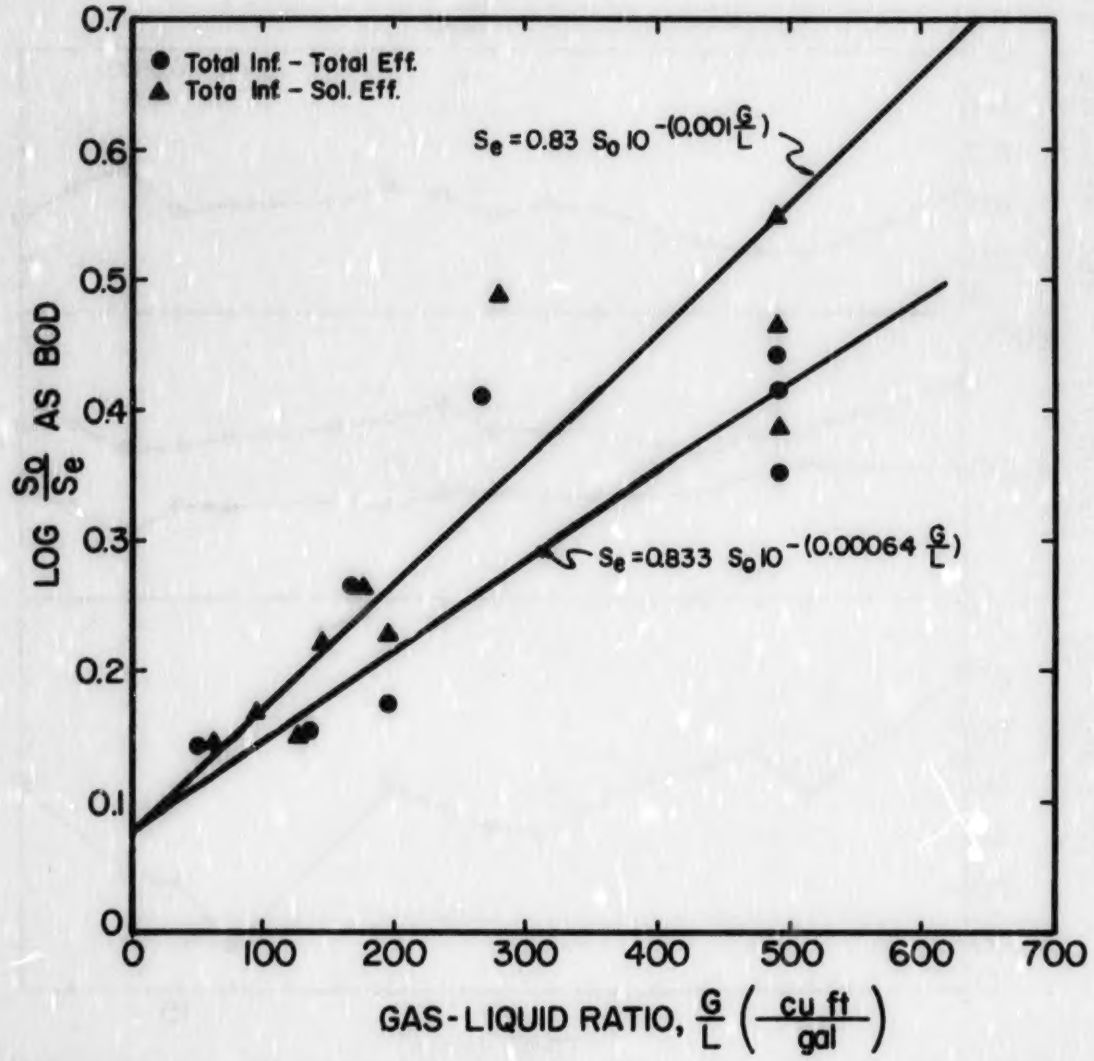


FIG.53. LINEARIZED DATA FOR PACKED TOWER STRIPPING - BOD BASIS
(cu m/cu m = cu ft/gal x 7.48)

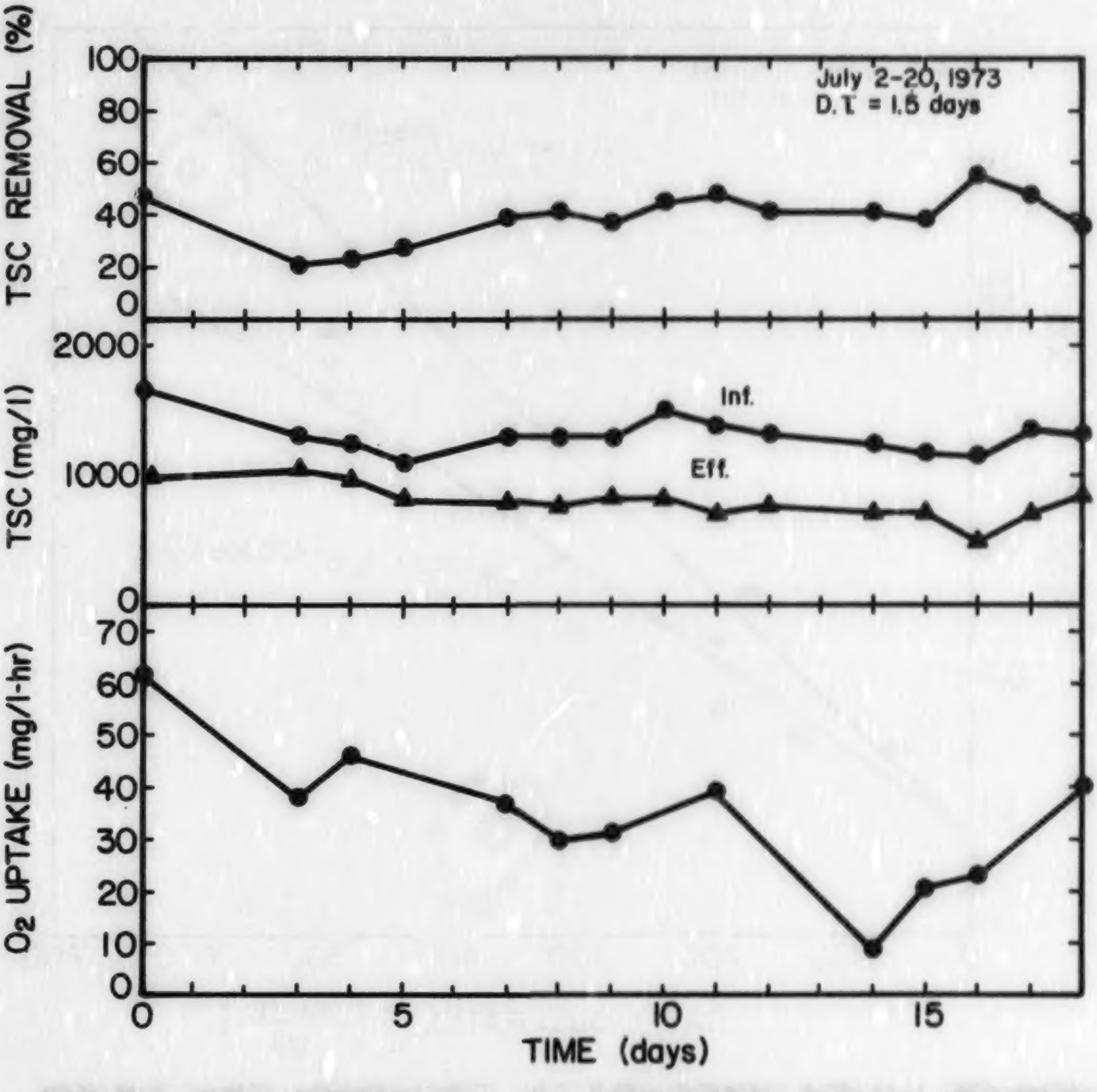


FIG.54. TSC REMOVAL IN AERATED EQUALIZATION BASIN – D.T. = 1.5 DAYS

TABLE 12

SUMMARY FOR STRIPPING IN AERATED BASIN

Detention time (days	Temp		Flow (gpm)[a]	TSC(mg/l)		TSC Removed[b] (%)	BOD (mg/l)				BOD removed[b] (%)	COD (mg/l)				COD removed[b] (%)	SS (mg/l)	
							Total		Soluble			Total		Soluble				
	Air	Basin		Inf.	Eff.		Inf.	Eff.	Inf.	Eff.		Inf.	Eff.	Inf.	Eff.		Inf.	Eff.
3.1	23	23	3	1579	548	65.3	6000[d]	2110[d]	4758	1453	59.8	6361	1990	6100	1400	68.7	211	409
3.0[c]	25	24	3	1264	522	58.7	3565	617	3685	1435	82.7	6240[c]	2029	--	1740	53.2	228	344
1.5	26	25	12	1327	804	39.4	4467	2009	4097	1680	55.0	5630	3128	4853	2473	44.4	168	321

[a] Flow in cu/min = gpm 3.79 x 10^{-3}

[b] Percent removals based on total influent and effluent values.

[c] During this period, 20 mg/l Cu^{++} was added to basin influent.

[d] Based on one value.

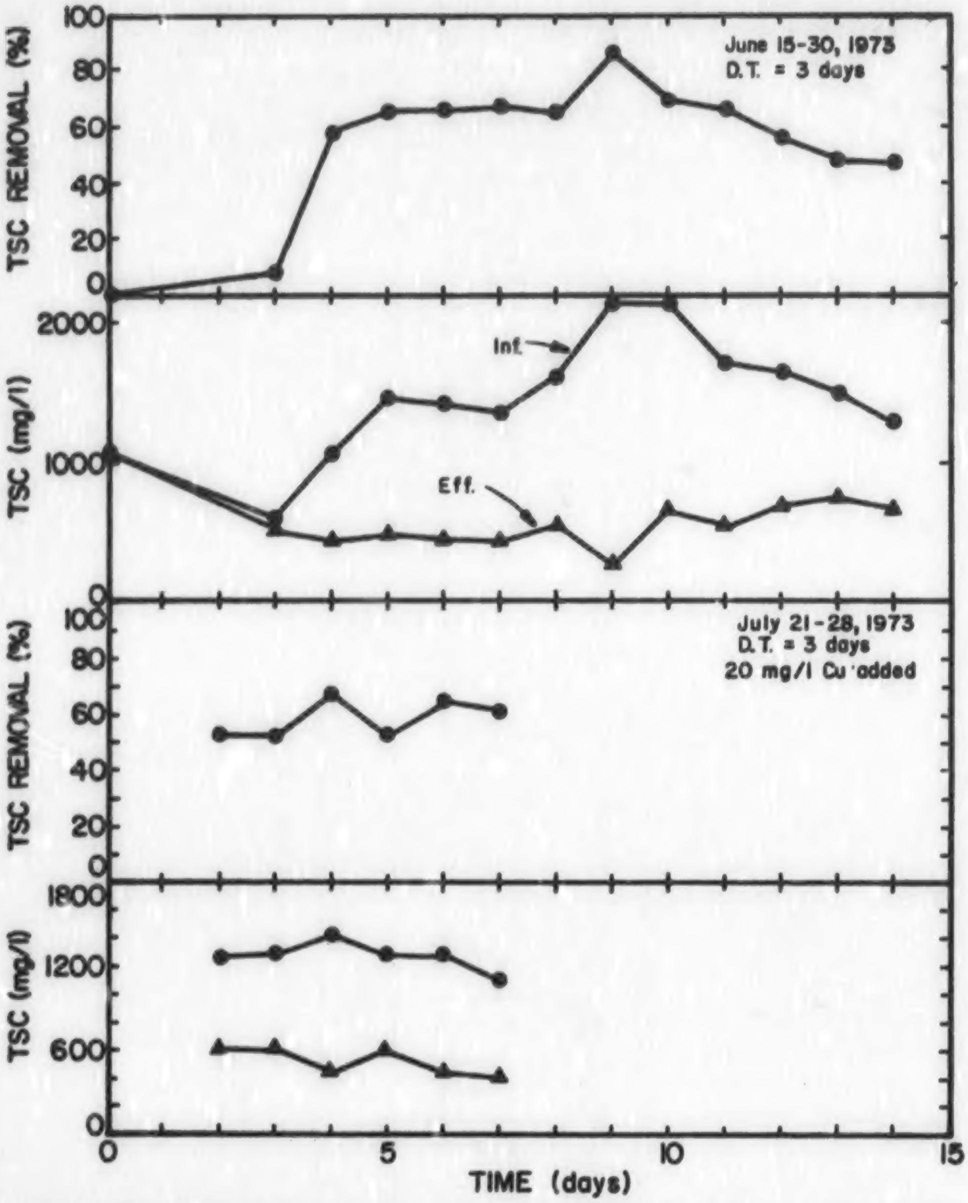


FIG. 55. TSC REMOVAL IN AERATED EQUALIZATION BASIN - D.T. = 3 DAYS

and 24 mg/l-hr for this 7-day test period. These measurements indicated that biological activity was significantly lower during this test period, although the value of 24 mg/l-hr indicates that complete inhibition of microorganisms was not achieved. Corresponding total carbon removal during these two tests were 59 and 65 percent, respectively, for tests with and without the copper addition indicating that only slightly reduced removals resulted during the tests in which copper was added to the system. This might be due to a complete inhibition of the microorganisms or to competition between the stripping and biological removal mechanisms. In the latter case, a greater removal of organic constituents due to air stripping might be realized under conditions where there is no removal by biological mechanisms.

Removals of BOD, COD, and TSC observed in these investigations is shown in Figure 56. The BOD removal for the 3-day detention time was significantly less than that which would be predicted from corresponding removals of COD and TSC. The expected removals of BOD have also been indicated in Figure 56. Because only one valid BOD determination was obtained from this test, it is felt that BOD removals of approximately 80 percent would be observed on an average basis under these conditions.

Removal of Specific Organic Constituents

Periodically during the continuous-flow stripping investigations, samples were taken and analyzed for solvents. All samples taken for solvent analysis were made on a grab basis. A summary of these results appears in Table 13. These data show very significant removals of toluene and isopropanol in both the aerated basin and the packed tower. Toluene was completely removed in the stripping processes for each observation.

Acetone concentrations increased in the aerated basin except for the sample taken on July 26 during the test in which copper was added to the basin influent. Decreases in acetone concentration were observed in the packed tower on all occasions except for the the sample taken on July 9. Data taken for the packed stripping tower followed by a biological trickling filter on July 26 showed that acetone decreased through the stripping tower but increased to its original level in the trickling filter. These data tend to corroborate an observation made in the trickling filter system concerning increases in acetone concentration. Acetone concentrations decreased in the packed tower and in the aerated basin

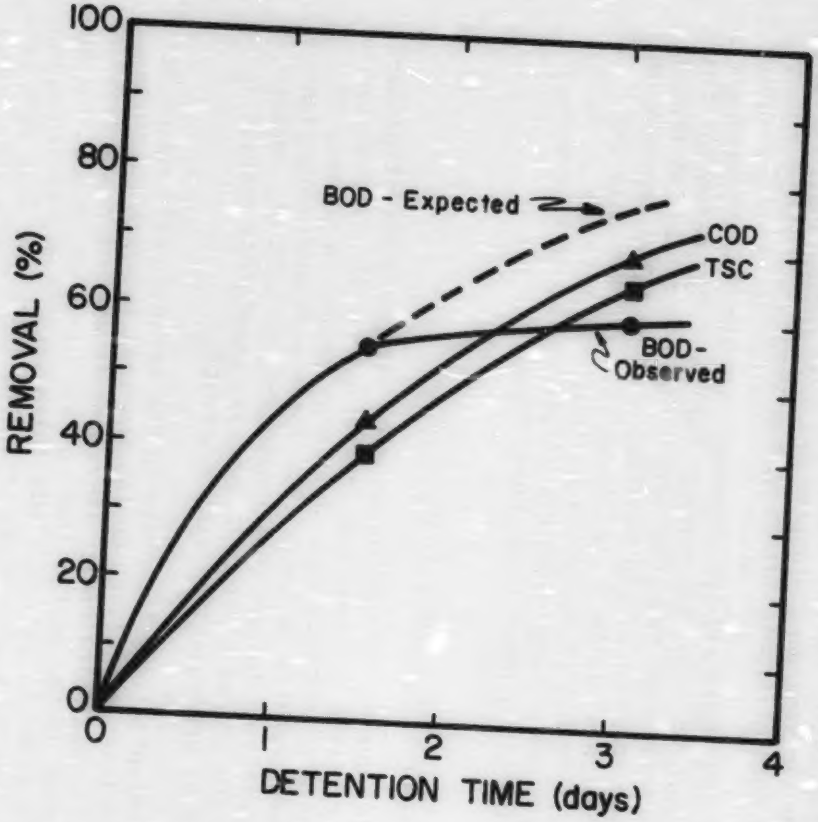


FIG. 56 .ORGANIC REMOVALS IN THE AERATED BASIN

TABLE 13

SOLVENTS MEASURED IN STRIPPING INVESTIGATIONS

Process	Date (July 1973)	Influent (mg/l)				Effluent (mg/l)			
		Acetone	Methanol	Isopropanal	Toluene	Acetone	Methanol	Isopropanal	Toluene
Aerated basin	9	300	100	1900	100	700	Tr^{6}	100	-
	12	200	-	1500	300	500	-	300	-
	23	200	Tr	1300	Tr	400	200	100	-
	26	400	400	1600	200	200	-	200	-
Packed tower	9	300	100	1800	100	700	100	200	-
	12	200	-	1500	Tr	100	-	700	-
	23	200	Tr	1300	200	100	-	400	-
	26	400	400	1600	-	300	200	800	-
Trickling filter Following packed tower	26	300	200	800	-	400	200	600	-

[a] Tr = Trace.

tests in which biological growths were inhibited. On the other hand, acetone concentrations increased in the aerated basin before copper was added and in the trickling filter following the packed stripping tower. This indicated that increases in acetone concentrations are due to by-products of biological degradation and not to autooxidation of isopropanol in the presence of air.

SECTION XII

ACKNOWLEDGEMENTS

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SECTION XIV

LIST OF SYMBOLS AND ABBREVIATIONS

a	Sludge yield coefficient (g VSS produced/g substrate removed)
a'	Assimilative oxygen utilization coefficient (g O_2/g substrate removed)
b	Endogenous respiration rate (g VSS destroyed/day-g MLVSS)
b'	Endogenous oxygen utilization coefficient (g O_2/g MLVSS-day)
BV	Bed volume; equal to the gross volume occupied by a material
C	Concentration (mg/l)
C_o	Influent concentration (mg/l)
D	Filter depth (ft)
D.O.	Dissolved oxygen
F/M	Organic loading, food to microorganism ratio (g substrate applied/g MLVSS-day)
G	Gas (air) flow (cfm) (cu m/min)
k	Activated sludge substrate removal rate constant (l/mg-day)
K	Trickling filter substrate removal rate constant ([gpm/sq ft]^n/ft) (cu m/min-sq m)^n/m)
L	Liquid (waste) flow (gpm) (cu m/min)
MLSS	Mixed liquor suspended solids
MLVSS	Mixed liquor volatile suspended solids
n	Flow exponent

N	Recycle ratio, $N = R/Q$
Q	Flowrate (liters/day or mgd) (cu m/day)
Q_A	Applied hydraulic loading (gpm/sq ft) (cu m/min-sq m)
R	Recycle flow (liters/day or mgd) (cu m/day)
R_r	Oxygen utilization rate (g O_2/day-1)
SS	Suspended solids
S_e	Effluent substrate concentration (mg/l)
S_o	Influent substrate concentration (mg/l)
S_r	Substrate removed, $S_r = S_o - S_e$ (mg/l)
t	Aeration time (days)
T, T'	Temperature (°C)
TDS	Total dissolved solids (mg/l)
TSC	Total soluble carbon (mg/l)
V	Volume (liters or mil gal)
VSS	Volatile suspended solids
X_v	Mixed liquor volatile suspended solids (mg/l)
ΔX_v	Excess volatile suspended solids (lb/day) (g/day)

SECTION XV

APPENDICES

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APPENDIX A

TREATMENT FACILITY COSTS

TABLE A1

TREATMENT FACILITY CAPITAL COST SUMMARY[a]

Item	Cost
Raw waste sump pumps, motors, and bar screen	$ 48,800
Solvent separation tank (50,000 gal, steel)	42,900
Neutralization (vessel, acid and caustic metering pumps, acid and caustic storage tanks)	17,400
Nutrient feed (phosphoric acid and anhydrous ammonia tank)	4,900
Equalization tanks, two (100 ft dia, 20 ft SWD, steel)	416,000
Pumping through preliminary treatment (pumps and motor)	26,400
Trickling filters, six (including packing)	540,000
Trickling filter fans and motors, six	6,600
Trickling filter pumps and motors, six	9,600
Clarifier Shell (50 ft dia, 10 ft SWD, steel)	52,000
Clarifier Rake	39,000
Sludge Pump	1,000
Instruments (including control panel)	32,900
Control building (three offices, conference room, laboratory, maintenance area, motor control room, equipment and chemical storage room)	165,000
Site work (including excavation, soil testing, and piping)	170,000
Engineering	277,500
Total	$1,850,000

[a] Costs are at 1970 levels.

TABLE A2

TREATMENT FACILITY OPERATING COST SUMMARY[a]

Item	Cost
Supervisory and administrative labor (including engineering support)	$ 26,000
Operating labor	67,000
Electricity	17,000
Chemicals	83,000
Laboratory and analytical costs	26,000
Maintenance (labor and materials)	83,000
Yardwork and custodial costs	24,000
Water	1,000
Total	$327,000/yr

[a] Costs are for calendar 1973.

APPENDIX B

TRICKLING FILTER DATA SUMMARY

TABLE B1

TRICKLING FILTER WEEKLY DATA SUMMARY

	BOD(mg/1)				COD (mg/1)				TSC(mg/1)			Temperature (°F)[a]			Flow (mgd)[b]	Loading (lb/1000 cu ft-day)[c]		Total Solids (mg/1)		SS (mg/1)		VSS (mg/1)	
	Inf	Eff		%	Inf	Eff		%	Inf	Eff	%												
Date	Total	Total	Sol	REM	Total	Total	Sol	REM	Total	Sol	REM	Inf	Eff	Air		BOD	COD	Inf	Eff	Inf	Eff	Inf	Eff
11/1	--	--	--	--	4090	--	2955	--	--	--	--	--	--	--	1.22	--	694	2614	2573	130	62	--	56
11/12	--	--	--	--	5048	--	3322	--	--	--	--	66	52	46	1.28	--	898	3114	2828	302	37	--	32
11/19	--	--	--	--	3930	2680	1947	31.8	1050	570	45.7	63	61	40	0.96	--	524	2843	2943	104	104	73	63
11/26	3300	--	--	--	3737	2953	2737	21.0	1151	934	18.9	63	58	41	1.07	491	556	328	694	96	71	86	63
12/3	4313	2848	2737	34.0	6030	4237	3538	29.7	1716	1291	24.8	63	58	36	1.28	767	734	--	--	80	99	63	80
12/10	4097	3177	3253	--	6100	4603	4153	24.5	1763	1513	14.2	63	58	33	1.18	672	1000	2540	3000	187	97	160	79
12/17	5985	5060	4200	15.4	5773	4730	3493	18.1	1732	1390	19.2	60	58	33	0.85	707	682	2650	2710	167	126	145	110
12/24	4545	3940	3365	13.3	4957	3175	2750	35.9	1016	680	33.1	61	57	34	0.90	569	620	3680	2360	225	69	182	67
12/31	4820	3280	2770	32.0	5970	4300	3805	28.0	1584	1233	22.2	58	56	37	1.05	703	871	2130	2260	98	83	89	79
1/7	3620	2630	1750	27.3	6753	5480	4587	18.8	1732	1495	13.7	56	50	14	1.15	579	1079	2125	2113	497	272	486	225
1/14	3460	3027	2297	12.5	5233	4000	3653	23.6	1543	1155	25.1	59	57	40	1.28	616	931	3230	9828	262	143	245	130
1/21	2570	1797	1607	30.1	4390	3347	2930	23.8	1180	898	23.9	60	57	38	1.25	446	763	--	--	141	129	116	116
1/28	2470	2335	2155	5.5	4453	3637	3240	18.3	976	916	6.1	59	56	26	1.21	415	749	4540	2373	119	120	111	109
2/4	2997	2480	2225	17.2	4707	3120	2727	35.7	1122	912	16.7	61	55	31	1.01	421	661	2210	2490	100	159	86	130
2/11	4337	2893	2620	33.3	5443	3910	3470	28.2	1348	1018	24.5	57	52	30	1.01	609	764	2970	2730	104	113	93	97
2/18	3080	2280	1950	26.0	4680	3540	2980	24.4	2950	2750	6.8	59	54	35	0.72	308	468	2950	2750	159	105	106	79
2/25	3558	2625	2805	26.2	5107	3947	3840	22.7	1495	1338	10.5	59	53	30	0.72	356	511	3012	2114	103	91	90	75
3/4	2672	1850	1707	30.8	4440	2580	1825	41.9	1337	847	36.6	61	59	45	0.50	186	309	2078	1884	95	84	74	71
3/11	4052	2347	2450	42.1	5773	3503	3147	39.3	1614	1013	37.2	62	62	49	0.50	230	401	3279	3645	101	205	86	161
3/18	2273	1281	1118	43.6	3507	2387	2126	31.9	1099	744	32.3	61	57	36	0.50	158	244	2639	2431	72	134	69	124
3/25	3035	1447	1141	52.3	4080	2547	2267	37.6	1037	668	35.6	63	63	41	0.72	304	409	--	2225	170	96	111	86
4/1	1987	1179	1056	40.7	3813	2360	2200	38.1	888	629	29.2	63	60	44	0.72	199	382	4860	2560	124	137	102	106
4/8	1258	629	414	50.0	2647	1485	1127	43.9	807	500	38.0	64	59	39	0.72	126	265	2430	2380	106	126	88	104
4/15	1793	888	582	50.5	3347	1747	1360	47.8	892	547	38.7	67	65	58	1.08	269	502	2390	2780	126	187	99	163
4/22	1164	464	355	50.1	2200	1187	787	46.0	727	296	59.3	67	66	57	1.03	167	315	2750	2930	78	232	75	209
4/29	1870	1323	1150	29.2	2600	1928	1784	25.8	748	545	27.1	68	63	54	0.99	142	358	2520	2140	78	129	66	110
5/6	1315	611	569	53.5	2960	2053	1720	30.6	940	695	26.1	69	63	54	0.94	172	387	2388	2133	89	105	80	94
5/13	1750	1031	1186	41.1	2947	2320	2107	21.3	842	691	17.9	70	62	54	0.99	241	405	2032	2652	111	132	105	121
5/20	1747	1223	1160	30.0	2960	2507	2403	15.3	867	729	14.9	70	63	55	1.10	267	453	2436	2743	98	89	92	86
5/27	1995	1810	1510	9.3	2890	2320	2160	19.4	659	660	0	71	69	67	0.91	252	364	--	--	127	99	124	97
6/3	2398	1753	1823	26.9	3253	2680	2367	17.6	786	689	12.3	75	72	70	1.15	383	520	2144	2244	220	152	196	142
6/10	2598	2292	2263	11.8	3371	3133	2963	7.1	800	785	1.9	78	77	77	0.91	329	426	1817	2297	193	155	150	131
6/17	4750	3700	3800	22.1	6253	4067	3813	35.0	1312	841	35.9	81	73	70	0.81	534	869	1497	2013	205	121	187	107
6/24	5400	--	5600	--	8240	5600	5160	32.0	1865	1408	24.5	80	75	70	0.94	705	1145	3694	3505	239	122	160	89
7/1	5227	3153	2930	39.7	7147	4707	4320	34.1	1644	1161	29.4	73	82	78	0.97	704	993	--	--	179	156	135	137
7/8	3963	2140	2030	46.0	6267	3573	3367	43.0	1365	1007	26.2	75	81	79	1.13	622	871	3590	3650	213	143	194	131
7/15	4037	2563	1760	36.5	6160	3893	3520	36.8	1285	991	22.9	72	82	77	1.10	617	856	3020	2770	235	164	200	143
7/22	4060	2170	2500	46.6	5760	3353	3107	41.8	1253	912	27.2	74	81	78	0.79	446	801	3420	3670	235	145	188	121

[a] Temperature in °C = 5/9 (F-32).

[b] Flow in cu m/day = mgd x 4,000.

[c] Loading in kg/cu m-day = lb/1,000 cu ft-day x 0.016.

TABLE B2

FULL-SCALE TRICKLING FILTER RECYCLE DATA

Date (1974)	Avg. Air Temp. (°C)	Inf. BOD (mg/l)	Bank A[a] Eff. BOD (mg/l)	Bank A[a] Removal (%)	Bank B[b] Eff. BOD (mg/l)	Bank B[b] Removal (%)
Apr. 1	9.4	1,546	1,435	7.2	892	42.3
Apr. 8	6.7	1,492	890	40.3	630	57.8
Apr. 15	11.7	1,522	1,453	4.5	1,030	32.3
Apr. 22	11.1	2,532	1,810	28.5	1,537	39.3
Apr. 29	13.3	2,351	1,836	21.9	1,382	41.2
May 6	10.6	2,851	2,555	10.4	1,696	40.5
May 13	17.8	2,720	2,097	22.9	1,472	45.9

[a] Operated at a recycle ratio of 0.43.

[b] Operated at a recycle ratio of 2.33.

APPENDIX C

ACTIVATED SLUDGE DATA

TABLE C1

SUMMARY OF ACTIVATED SLUDGE RESULTS-PHASE I

UNIT	IA[a]	IB[b]	IC[b]	ID[b]	IE[c]
OPERATIONAL PARAMETERS					
F/M (g BOD/day-g MLVSS)	0.22	0.084	0.23	0.40	0.24
Detention time (days)	5.3	11.3	4.1	2.1	3.6
MLVSS (mg/l)	2620	2410	2440	2700	2600
Volume (l)	151	151	151	151	151
D.O. (mg/l)	6.2	6.7	5.9	4.9	5.8
Oxygen uptake (g O_2/day-g MLVSS)	0.22	0.11	0.23	0.45	0.26
SVI (ml/g)	36	48	39	33	40
ZSV (ft/hr) (m/m)	14.7	14.4	24.1	19.5	13.6
Temperature (°C)	17	17	17	17	17
Sludge age (days)	15.5	32.9	15.3	7.5	13.5
INFLUENT QUALITY PARAMETERS					
BOD (mg/l)					
Total	3070	2270	2270	2270	2220
Soluble	2910	2200	2200	2200	2160
COD (mg/l)					
Total	4760	3860	3860	3860	3900
Soluble	3980	3500	3500	3500	3510
pH (median value)	7.3	7.3	7.3	7.3	7.1
EFFLUENT QUALITY PARAMETERS					
BOD (mg/l)					
Total	77	15	53	233	39
Soluble	54	14	47	184	39
COD (mg/l)					
Total	723	416	603	825	547
Soluble	421	336	384	519	379
pH (median value)	6.1	7.2	6.6	5.9	6.3
SS (mg/l, median value)	300	100	214	302	182
ORGANIC REMOVAL EFFICIENCIES					
BOD (%)	98.2	99.4	97.3	91.9	98.2
COD (%)	91.2	91.3	90.0	86.5	90.3
BOD removal rate (g BOD_r/day-g MLVSS)	0.218	0.083	0.221	0.36	0.232
COD removal rate (g COD_r/day-g MLVSS)	0.314	0.130	0.347	0.581	0.376

[a] Unit fed raw equalized waste.
[b] Unit fed unclarified trickling filter effluent.
[c] Unit fed clarified trickling filter effluent.

TABLE C2

SUMMARY OF ACTIVATED SLUDGE RESULTS - PHASE II[a]

Influent	Unclarified Trickling Filter Effluent		
Unit	**IIC**	**IID**	**IIF**
OPERATIONAL PARAMETERS			
F/M (g BOD/day-g MLVSS)	0.23	0.43	0.23
Detention time (days)	2.18	1.19	2.48
MLVSS (mg/l)	3150	3020	2760
Reactor volume (l)	151	151	151
D. O. (mg/l)	6.9	6.4	6.7
Oxygen uptake (g O_2/g MLVSS-day)	0.28	0.45	0.31
Sludge age (days)	16.7	11.4	22.7
Temp (°C)	18	18	14
INFLUENT QUALITY PARAMETERS			
BOD, Total (mg/l)	1559	1559	1559
COD, Total (mg/l)	2096	2096	2172
Soluble (mg/l)	1610	1630	-
pH (median value)	6.9	6.9	7.0
EFFLUENT QUALITY PARAMETERS			
BOD (mg/l)			
Soluble	47	111	44
COD (mg/l)			
Total	365	516	354
Soluble	261	350	268
pH (median value)	7.0	6.9	7.0
ORGANIC REMOVAL EFFICIENCIES			
BOD (%)	97.8	94.4	96.8
COD (%)	87.5	83.3	87.7
BOD removal rate (g BOD_r/g MLVSS-day)	0.22	0.40	0.22
COD removal rate (g COD_r/g MLVSS-day)	0.27	0.48	0.28

[a] Includes data collected from March 10 to May 26, 1973 except for the cold temperature operating period from March 24 through April 13 excluded from Unit II F results.

TABLE C3

SUMMARY OF ACTIVATED SLUDGE RESULTS, VARIED TEMPERATURE OPERATION - PHASE II[a]

Influent	Unclarified Trickling Filter Effluent		
Unit	IIC	IID	IIF
OPERATIONAL PARAMETERS			
F/M (g BOD/day-g MLVSS)	0.130	0.240	0.142
Detention time (days)	2.60	1.50	4.44
MLVSS (mg/1)	3260	3070	1750
Reactor volume (1)	151	151	151
D. O. (mg/1)	6.7	6.6	9.3
Oxygen uptake (g O_2/g MLVSSOday)	0.28	0.31	0.51
Sludge age (days)	32.3	35.7	48.8
Temp (°C)	17	17	6
INFLUENT QUALITY PARAMETERS			
BOD, Total (mg/1)	1107	1107	1107
COD, Total (mg/1)	2546	2546	2450
Soluble (mg/1)	2165	2161	2027
pH (median value)	6.8	6.8	6.8
EFFLUENT QUALITY PARAMETERS			
BOD (mg/1)			
Soluble	26	49	71
COD (mg/1)			
Total	416	666	406
Soluble	273	375	301
pH (median value)	7.0	7.0	6.8
ORGANIC REMOVAL EFFICIENCIES			
BOD (%)	97.7	95.6	93.6
COD (%)	89.3	85.3	87.7
BOD removal rate (g BOD_r/g MLVSS-day)	0.128	0.230	0.133
COD removal rate (g COD_r/g MLVSS-day)	0.268	0.471	0.277

[a] Includes data collected from March 24 to April 13, 1973.

TABLE C4

SUMMARY OF ACTIVATED SLUDGE RESULTS-PHASE III[a]

Feed[b]	Unstripped		Stripped	
Aerating Gas	Air	Oxygen	Air	Oxygen
UNIT	III A	III B	III C	III D
OPERATIONAL PARAMETERS				
F/M (g BOD/day-g MLVSS)	0.31	0.34	0.29	0.32
Detention time (days)	5.2	3.43	2.4	2.3
MLVSS (mg/l)	2540	3500	2470	2210
Volume (l)	151	151	151	151
D.O. (mg/l)	4.1	8.8	4.1	8.6
Oxygen uptake (g O_2/day-g MLVSS)[c]	0.35	0.46	0.25	0.37
SVI (ml/g)	55	37	61	47
ZSV (m/hr)	17.1	15.3	14.1	18.0
Temperature (°C)	23	25	23	25
Sludge age (days)	12.4	14.1	17.1	28.0
INFLUENT QUALITY PARAMETERS				
BOD (mg/l)				
Total	4060	4030	1710	1620
Soluble	3720	3390	1440	1430
COD (mg/l)				
Total	5737	5440	2484	2410
Soluble	4972	4806	2016	1939
pH (median value)	8.2	8.3	8.3	8.3
EFFLUENT QUALITY PARAMETERS				
BOD (mg/l)				
Total	134	521	82	51
Soluble	99	366	72	47
COD (mg/l)				
Total	836	989	373	321
Soluble	432	546	320	276
pH (median value)	7.8	7.1	7.9	7.7
SS (mg/l, median value)	240	280	130	115
ORGANIC REMOVAL EFFICIENCIES				
BOD (%)	97.6	89.5	96.3	97.4
COD (%)	92.5	90.0	87.1	88.5
BOD removal rate (g BOD_r/day-g MLVSS)	0.30	0.31	0.28	0.31
COD removal rate (g COD_r/day-g MLVSS)	0.40	0.41	0.37	0.42

[a] Includes data from June 18 to July 28, 1973.
[b] All units fed raw equalized waste with or without prestripping as indicated.
[c] First week of data not included in averages.